

DOE/NETL-2001/1156

Milliken Clean Coal Demonstration Project: A DOE Assessment

August 2001

U.S. Department of Energy
National Energy Technology Laboratory

P.O. Box 880, 3610 Collins Ferry Road
Morgantown, WV 26507-0880

and

P.O. Box 10940, 626 Cochrans Mill Road
Pittsburgh, PA 15236-0940



Disclaimer

This report was prepared as an account of work sponsored by an agency of the United States Government. Neither the United States Government nor any agency thereof, nor any of their employees, makes any warranty, express or implied, or assumes any legal liability or responsibility for the accuracy, completeness, or usefulness of any information, apparatus, product, or process disclosed, or represents that its use would not infringe privately owned rights. Reference therein to any specific commercial product, process, or service by trade name, trademark, manufacturer, or otherwise does not necessarily constitute or imply its endorsement, recommendation, or favoring by the United States Government or any agency thereof. The views and opinions of authors expressed therein do not necessarily state or reflect those of the United States Government or any agency thereof.

Contents

	Page
Executive Summary	5
I Introduction.....	9
II Project/Process Description	10
II.A Project Description.....	10
II.B Need for the Technology Demonstration.....	10
II.C Promise of the Technology	11
II.D Technology Description.....	13
II.E Project Implementation.....	17
II.F Project Objectives and Statement of Work.....	21
III Technical and Environmental Assessment	23
III.A Technical Results	23
III.B Environmental Performance	28
III.C Impacts on Equipment	30
III.D Evaluation of Project Achievements	32
IV Market Analysis	33
IV.A Market Size and Penetration.....	33
IV.B Economics.....	33
V Conclusions.....	36
VI Abbreviations.....	38
VI References.....	39
VII Bibliography.....	40

List of Tables and Figures

Table		Page
1	Properties of Design Coals From Public Design	24
2	Spray Header Results	26
3	S-H-U Retrofit Capital Cost.....	34
4	S-H-U Operating Costs	35
5	S-H-U Costs for Sulfur-Dioxide Removal.....	35

Figure		Page
1	Block Diagram for the MCCTD Project	42
2	S-H-U Flow Diagram.....	42
3	S-H-U FGD Absorber	43
4	Equilibrium Diagram for the $H_2SO_3 - HSO_3^- - SO_3^{2-}$ Systems	43
5	pH as a Function of Position in S-H-U Cocurrent/Countercurrent Absorber	44
6a	Typical Heat Pipe Air Heater Construction	44
6b	Heat Pipe Cross Section.....	45
7	Limestone Preparation Flow Diagram	45
8	By-Product DeWatering Flow Diagram.....	46
9	Blowdown Treatment/Brine Concentration Flow Diagram.....	46
10	Schematic Showing Approximate Locations of Solid Deposits in Absorber	47
11	SO ₂ Removal vs. Total Number of Headers in Operation	47
12	SO ₂ Removal vs. Liquid/Gas Ratio	48
13	SO ₂ Removal vs. Formic Acid Concentration	48
14	Effect of CaCl ₂ Concentration on SO ₂ Absorption Efficiency	49
15	Effect of Limestone Grind on SO ₂ Removal	49

Executive Summary

This document is a U.S. Department of Energy (DOE) post-project assessment of the Milliken Clean Coal Technology Demonstration (MCCTD) Project funded in Round IV of the Clean Coal Technology (CCT) Program. In October 1992, the New York State Gas & Electric Corporation (NYSEG) entered into a cooperative agreement with DOE to conduct this study and to jointly fund the project, which was sited at NYSEG's Milliken Station located at Lansing, NY, on the East shore of Cayuga Lake. The MCCTD Project consisted of installing nitrogen oxides (NO_x)- and SO₂ (sulfur dioxide)-reduction technologies on units 1 and 2 at Milliken Station. Each unit has a rated capacity of 150 MWe, for a total station capacity of 300 MWe. DOE provided 28 percent of the total project funding of \$159 million. Construction for the demonstration project began in April 1993 and ended in June 1995; operations were initiated in January 1995 and completed in November 1998. Operating over 30,000 hours during 3½ years, this project was very successful, yielding substantial environmental benefits in the form of emissions and waste reductions. Its operational success is validated by the continued use of equipment to lower emissions from the Milliken Station.

In addition to NYSEG, the Empire State Electric Energy Research Corporation, Consolidation Coal Company (CONSOL), the Electric Power Research Institute (EPRI), and the New York State Energy Research and Development Authority (NYSERDA) were cofunders. Others participating in the project were Saarberg-Hölter-Umwelttechnik (S-H-U), GmbH, The Stebbins Engineering and Manufacturing Company, ABB Air Preheater, Inc., and DHR Technologies, Inc., technology suppliers. Gilbert/Commonwealth (G/C) provided design and construction management services.

The heart of the MCCTD Project is the S-H-U Flue Gas Desulfurization (FGD) technology. Although the S-H-U process was developed and tested in Europe, this project represents the first test on a boiler burning U.S. coals, which have significantly different properties and trace contaminants than European coals. This demonstration is a critical step before the innovative S-H-U technology can make a significant contribution to decreasing SO₂ emissions in the U.S. Furthermore, the S-H-U installation at Milliken incorporated some new design features that had not previously been demonstrated and which significantly improve the attractiveness of the process. These include a split-flow design for the absorber, use of corrosion-resistant Stebbins tile to line the absorber, and space-saving placement of the absorber directly below the stack.

Other features of the MCCTD Project include the Low-NO_x Concentric Firing System Level III (LNCFSTM III) to control NO_x, a high-efficiency heat pipe air preheater to use energy efficiently, gypsum production and sales to eliminate land filling of waste solids, and production of a salable concentrated brine to help eliminate water pollution. In addition, the Plant Emission Optimization Advisor (PEOATM) provided state-of-the-art artificial-intelligence-based control of key plant operating parameters. MCCTD was designed as a zero-discharge project with no net increase in power consumption even with added environmental controls.

In the S-H-U process, SO₂ is absorbed in a limestone slurry stored in a novel absorber with both cocurrent and countercurrent slurry-spray zones. When SO₂ is adsorbed in water, it forms

bisulfite ions. In the S-H-U process, small amounts of formic acid are added to buffer the slurry and keep it in the pH range of 4-5 throughout the entire spray zone. This pH range ensures that water soluble calcium bisulfite is formed. For all load conditions, the S-H-U process with its buffered slurry operates within the pH range that precludes sulfite formation, thus avoiding scaling and plugging in the absorber, and greatly reducing maintenance requirements compared to unbuffered processes. The ability to operate in a nonscaling mode, even during transients, may be the biggest advantage to low pH-buffered absorption.

Flue gas from the electrostatic precipitators (ESP) flows to the top of the cocurrent section of the absorber, where it is first cooled and saturated by a water spray and then contacted with recycle slurry from spray nozzles at up to four levels. At the bottom of the cocurrent zone, the slurry droplets disengage from the flue gas and collect in the absorber reaction tank, while the flue gas reverses flow and passes to the countercurrent section, where it is contacted with recycle slurry from spray nozzles at up to three levels to complete SO₂ absorption. After passing through a two-stage mist eliminator, the flue gas is discharged to a wet stack, which sits directly above the absorber. The Milliken project provided the first demonstration of the S-H-U process installed directly beneath a plant stack, a design approach that saves considerable space. The system is designed to remove 95 percent of the SO₂ when burning 3.2-percent sulfur coal with five of the seven spray levels in operation, with high-chloride levels in the recycle slurry.

An additional feature of the MCCTD Project is a tile-lined, split-module absorber design provided by The Stebbins Engineering and Manufacturing Company. The absorber is a concrete vessel with tile lining and a common center dividing wall to provide each unit with its own absorber module. Each side of the vessel operates independently of the other. The tile lining provides superior abrasion and corrosion resistance when compared to rubber and alloy linings and should last for the life of the plant. Also, in the event that a leak should occur, the tile lining allows it to be repaired while the unit is operating.

Limestone is the primary reagent for reaction with absorbed SO₂. The equipment for limestone grinding, fresh limestone slurry storage, and fresh slurry transfer to the S-H-U unit is located in the limestone preparation area of the FGD building. Limestone is continuously pumped from the fresh slurry feed tank to the absorbers.

The slurry droplets from both cocurrent and countercurrent absorber sections collect in the absorber reaction tank. The reaction tank acts as a back-mixed reactor to complete oxidation of calcium bisulfite to calcium sulfate. Air from blowers is bubbled through the reaction tank to oxidize the bisulfate ions to sulfate ions, which then react with calcium ions to form insoluble hydrated calcium sulfate (gypsum). The large gypsum crystals that form are easy to dewater and are attractive to wallboard manufacturers because they produce a high-quality wallboard. Agitators prevent gypsum particles from settling to the bottom of the tank.

A slurry slipstream is pumped from the absorber reaction tank to the dewatering area, where the slurry is concentrated to approximately 45-percent solids by hydrocyclones and further dewatered to greater than 90-percent solids by centrifuges. The gypsum cake is transported to a storage building, where it is stored until loaded onto trucks for delivery to the purchaser. A slipstream of the recovered water is sent to blowdown treatment and brine concentration. The rest of the water is either sent to the limestone grinding system or recycled to the absorber. The brine

concentration system was installed to recover a highly concentrated CaCl_2 brine that could be sold commercially. Unfortunately, problems that were not fully resolved by the end of the project prevented satisfactory operation of this unit.

Another feature of this project is an energy savings provided by a heat pipe (Q Pipe) air preheater, which was installed to replace the Ljungstrom regenerative type air preheater on Unit 2. The heat pipe design has the potential to eliminate many of the problems, such as air in-leakage and plugging, associated with tubular and Ljungstrom air preheater designs, and to operate at somewhat lower flue gas outlet temperatures, thus improving overall plant heat rates and station operation.

Combustion modifications, installed on both Milliken Units 1 and 2 for NO_x control, consisted of replacing the existing conventional tangential firing system with LNCFSTM III furnished by ABB Combustion Engineering. LNCFSTM III maximizes the NO_x -reduction capabilities of existing tangential firing systems while minimizing unit modifications. PEOATM, an on-line performance support system developed by DHR Technologies, Inc., to assist plant personnel in meeting environmental standards and in optimizing overall plant economic performance, was installed on both Milliken units 1 and 2.

LNCFSTM III lowered NO_x emissions from a baseline value of 0.64 to 0.39 pounds per million Btu (lb/Mbtu), a 39-percent reduction, at full boiler load and from 0.58 to 0.41 lb/MBtu (29 percent reduction) at reduced load. These results are within the expected range for low- NO_x burners. Boiler efficiency with LNCFSTM III was 88.3-88.5 percent compared to 89.3-89.6 percent at baseline, but when adjustments were made for differences in operating conditions and LNCFSTM III and the baseline were compared at similar flue gas temperatures and compositions, the estimated LNCFSTM III boiler efficiency was 0.2 percent higher than baseline. With LNCFSTM III, unburned carbon in fly ash stayed below 4 percent, and carbon monoxide (CO) emissions did not increase.

Tests of the S-H-U process were run burning low (1.6 percent), design (2.23 percent), and high-sulfur (3.2-4.1 percent) coals. These tests indicate that: (1) depending on operating conditions, SO_2 removal up to 98 percent can be achieved; (2) adding formic acid significantly enhanced SO_2 removal; (3) the pH of the spray slurry, the liquid-to-gas ratio, and the limestone grind size were important variables; and (4) formic acid increased the mass-transfer rate, but the effect diminished with increasing formic acid concentration.

The design-coal test was shortened because it was necessary to finish the project by the end of 1998. The high-sulfur coal test was also shortened because the limestone slurry feed pumps were unable to meet demand. Thus, the pH was not kept under control. Poor performance with high-sulfur coal was a result of trying to run above design conditions. If the S-H-U unit had been designed for high-sulfur coal, satisfactory operations could have been achieved.

The S-H-U process has wide applicability for SO_2 capture from flue gas within the utility and industrial sectors. With minor modifications, this process has been used in Europe to reduce SO_2 emissions from utility and industrial boilers firing coal, lignite, oil, and gas, as well as municipal-waste incinerators. The process is applicable to boilers firing low-, medium-, and high-sulfur coals regardless of boiler size or type and can achieve up to 98 percent SO_2 removal. All wet

limestone FGD processes require significant plot areas, but that required by S-H-U can be minimized by adoption of the split module, below-the-stack configuration demonstrated in the Milliken project.

The total capital requirement for an S-H-U FGD retrofit for a 300-MWe power plant to remove 95 percent of the SO₂ when burning 3.2-percent sulfur coal is approximately \$90 million (\$300/kW). However, when estimating costs for a particular installation scope, adjustments and site-specific factors need to be taken into account. For a 300-MWe power plant with a 65-percent operating factor, operating costs, including credit for gypsum sales at \$3/ton, are \$4.6 million/year.

Based on the above costs, an estimated levelized cost of 12 mills/kWh (\$533/ton of SO₂ recovered) on a current dollars basis or 9.2 mills/kWh (\$413/ton of SO₂ recovered) on a constant dollars basis. These costs appear to be competitive with regenerative flue gas scrubbing processes, although they are somewhat higher than other wet limestone scrubbing processes. However, economics is not the only factor that will promote the acceptance of the S-H-U process. Another factor is the highly beneficial environmental impact. Not only is the S-H-U process capable of SO₂ removals up to 98 percent, but instead of creating acres of sludge ponds, it produces a high-quality gypsum by-product, which wallboard manufacturers prefer over natural gypsum. Also, although not yet fully perfected, the prospect of producing a salable brine concentrate appears highly likely, thus eliminating another disposal problem.

To assess the potential market for the S-H-U process, the total U.S. electric market was divided into two parts: retrofit capacity (pre-New Source Performance Standards [NSPS] coal-fired boilers not equipped with SO₂ controls) and new capacity (projected coal-fired additions through 2030). This analysis indicates a potential total retrofit market of 5,700 MWe through 2030 and a new power plant market of 96,200 MWe. The successful demonstration of the S-H-U process at Milliken, along with the experience in Europe, should enable S-H-U to effectively market its FGD technology in the U.S. through its U.S. design and manufacturing partners.

I Introduction

The goal of the U.S. Department of Energy's (DOE) Clean Coal Technology (CCT) program is to furnish the energy marketplace with a number of advanced, more efficient, and environmentally responsible coal-utilization technologies through demonstration projects. These projects seek to establish the commercial feasibility of the most promising advanced coal technologies that have developed beyond the proof-of-concept stage.

This document serves as a DOE post-project assessment of a project selected in CCT Round IV, the Milliken Clean Coal Technology Demonstration (MCCTD) Project (see Figure 1), as described in a Report to Congress (U.S. Department of Energy, 1992). The need to meet strict emissions requirements with a minimum loss in generating efficiency prompted the New York State Electric & Gas Corporation (NYSEG) to submit the proposal for this project. In October 1992, NYSEG entered into a cooperative agreement with DOE to conduct the study. The project was sited at NYSEG's Milliken Station located on the East shore of Cayuga Lake in Lansing, New York, about 12 miles Northwest of Ithaca. The purpose of this CCT project was to demonstrate the reduction of SO₂ and NO_x emissions without a significant decrease in plant efficiency by installing a combination of innovative technologies and plant upgrades. These include the Saarberg-Hölter-Umwelttechnik (S-H-U) process for SO₂ reduction, the Low-NO_x Concentric Firing System Level III (LNCFSTM III) for NO_x reduction, and a high-efficiency heat pipe (Q Pipe) air preheater system plus other energy-saving modifications to maintain efficiency. DOE provided 28 percent of the total project funding of \$159 million.

The two balanced draft mode units at Milliken have a rated capacity of 150 MWe each, for a total station capacity of 300 MWe. The units are tangentially fired with four levels of burners in each of the four corners. Unit 1 was completed in 1955 and Unit 2 in 1958. The Milliken units are typical of many power plants in the U.S., and the demonstration of technologies at this scale is sufficient to establish their commercial viability. Based on data gathered in this demonstration, there should be no difficulty in scaling up to larger units.

Construction for the demonstration project was started in April 1993 and completed in June 1995. Operations were initiated in January 1995 and completed in November 1998. The independent evaluation contained herein is based primarily on information from NYSEG's Final Report (New York State Electric & Gas Corporation, 1999), as well as other references cited (New York State Electric & Gas Corporation, 1993; CONSOL, 1999a; CONSOL, 1999b; CONSOL, 2000).

II Project/Process Description

II.A Project Description

The major technology installed for the MCCTD Project was the S-H-U process for sulfur dioxide control, which treated the flue gas from both Units 1 and 2. Combustion modifications were also installed on both Units 1 and 2 and consisted of replacing the existing conventional tangential firing systems with LNCFS™ III, supplied by ABB Combustion Engineering. On each furnace LNCFS™ III included new burners, wind boxes, and an overfire air supply. In addition, a high-efficiency heat pipe air preheater was installed on Unit 2. The Plant Economic Optimization Advisor (PEOA™), an on-line performance-support system developed by DHR Technologies, Inc., was also installed on both units 1 and 2.

In addition to NYSEG, the Empire State Electric Energy Research Corporation, Consolidation Coal Company (CONSOL), the Electric Power Research Institute (EPRI), and the New York Energy Research and Development Authority (NYSERDA) were cofunders. Others participating in the project were Saarberg-Hölter-Umwelttechnik, GmbH, the Stebbins Engineering and Manufacturing Company, ABB Air Preheater, Inc., and DHR Technologies, Inc., who supplied the technologies. Gilbert/Commonwealth (G/C) provided design and construction management services.

II.B Need for the Technology Demonstration

Although the S-H-U Flue Gas Desulfurization (FGD) process was developed and tested in Europe, this project represents the first test on a boiler burning U.S. coals, which have significantly different properties and trace contaminants than European coals. Successful completion of this project is a critical step toward industry's acceptance of the S-H-U process, so that this innovative technology can make a significant contribution to decreasing SO₂ emissions in the U.S. Furthermore, the installation at Milliken incorporated some new design features that had not previously been demonstrated and which significantly improve the attractiveness of the technology. These include a split-flow design for the absorber, use of corrosion-resistant Stebbins tile to line the absorber, and space-saving placement of the absorber directly below the stack.

In addition to the S-H-U demonstration, this project included several other features designed to improve environmental performance and efficiency: LNCFS™ III for NO_x control, a high-efficiency heat pipe air preheater for energy efficiency, gypsum production and sales to eliminate land filling of solid wastes, brine concentration to help eliminate water pollution by producing a salable product, and PEOA™ to provide state-of-the-art artificial-intelligence-based control of key plant operating parameters to assist in meeting environmental standards and optimizing overall plant performance.

II.C Promise of the Technology

The promise of the technology installed under the MCCTD Project is to achieve large reductions in emissions of SO₂ and NO_x while experiencing at most a modest decrease in plant efficiency, maintaining plant operability, and preventing pollution by producing salable by-products, thus eliminating wastes. Another promise is better sorbent utilization, thus achieving both economic and environmental benefits. The promise of the individual technologies is discussed below.

II.C.1 S-H-U Process

Installation of the S-H-U technology was expected to achieve the following:

- Up to 98-percent SO₂ removal by operation of a two-stage cocurrent/countercurrent absorber
- Low limestone reagent consumption
- Excellent process stability and ease of operations during load changes and other operational transients
- Closed-cycle operation with essentially no liquid waste discharge
- Freedom from scaling and plugging in the absorber and associated equipment
- High absorber availability and low maintenance
- Production of commercially salable by-products (wallboard-grade gypsum and concentrated calcium chloride brine)
- Improved energy efficiency compared to other FGD technologies
- Ability to operate at high-chloride concentrations without adversely affecting SO₂-absorption efficiency

II.C.2 NO_x Reduction

As part of this project, LNCFSTM III, supplied by ABB Combustion Engineering was installed on both furnaces. The objective of this technology was to achieve up to 40-percent NO_x reduction at full load compared to the burners being replaced while maintaining an acceptable carbon-in-ash level. Originally, it was intended to install the NO_xOUTTM selective noncatalytic reduction (SNCR) technology in addition to low-NO_x burners, but for technical and economic reasons, this was deleted from the project; therefore, the NO_xOUTTM technology is not discussed in this report.

II.C.3 Heat Pipe Air Preheater

The heat pipe air preheater installed on Unit 2 represents the first time in the U.S. that a heat pipe has been used for both primary and secondary air preheating on a utility boiler. Potential benefits from using heat pipes for this service are:

- Reduced ID fan power requirements because air leakage, typical of conventional air preheaters, has been eliminated

- Reduced size of downstream environmental control equipment and reduced power requirements for the primary and secondary air fans because flue gas volume has been reduced as a result of less leakage
- Improved thermal efficiency because the heat pipe air preheaters can operate at a lower flue gas exit temperature
- Better control of corrosion in the heat pipe air preheater because of more stable cold-end operation and ease of cleaning with sootblowing and water washing, if required
- Full-shop fabrication

II.C.4 Other Technologies

Other technologies implemented to enhance plant performance include the Stebbins tile to line the absorber, DHR's PEOA™, the CAPCIS corrosion-monitoring system on the heat pipe, and a brine-concentration system.

Corrosion of the absorber is a problem that has plagued many FGD installations. Reinforced concrete/Stebbins ceramic tile promises to be a cost-effective alternative construction technique and offers the following potential advantages compared to conventional scrubber construction:

- Superior corrosion and abrasion resistance leading to longer absorber liner life and lower life-cycle costs
- High reliability and availability
- Suitability for construction at units with little free plot area
- Capability for on-line repairs and lower maintenance costs
- Ability to withstand higher temperatures and temperature excursions
- Ability to withstand high chloride levels, thus offering fuel flexibility

PEOA™ is an on-line support system developed by DHR Technologies to help meet economic performance targets by integrating key information and analyses that assist plant personnel in optimizing plant performance, including steam- and waste-management systems. PEOA™ automatically determines and displays key operational and control set points for optimized cost operation. Although not necessary for operation of the S-H-U technology, the promise of PEOA™ compensates for parasitic power losses and allows the Milliken Station to significantly reduce emissions while still maintaining high efficiency.

The purpose of the CAPCIS corrosion-monitoring system is to reduce corrosion on the heat pipe air preheater on which it was installed. Each air preheater has an air-side bypass duct for cold-end corrosion protection. The damper can be opened to allow combustion air to bypass the air preheater, thus reducing heat load and preventing the temperature of the flue gas from falling below the acid dew point. The CAPCIS system includes a number of sensors installed at strategic locations. The outputs from these sensors are sent to a programmable logic controller (PLC) and compared to predefined set points. The PLC then determines the proper setting for the bypass damper. The promise of this technology was to significantly reduce corrosion in the heat pipe air preheater while maintaining efficiency.

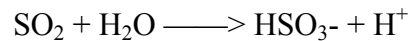
Chlorine in the coal ends up as calcium chloride in the scrubber slurry, and the CaCl_2 must be removed to prevent its buildup. Disposal of this material is a major environmental problem, and the MCCTD Project includes a system to produce concentrated calcium chloride brine, suitable for commercial sales. Thus, the promise of this technology was to eliminate a waste-disposal problem.

II.D Technology Description

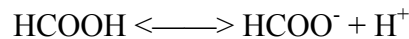
The following sections discuss the technologies installed as part of the MCCTD Project.

II.D.1 S-H-U Process

In the S-H-U process (see Figures 2 and 3), SO_2 is absorbed in a limestone slurry in a unique absorber with both cocurrent and countercurrent slurry spray zones. A novel feature of the S-H-U process is the addition of formic acid to the slurry to buffer the system and control the pH. When SO_2 is absorbed in water, it reacts to form bisulfite ions:



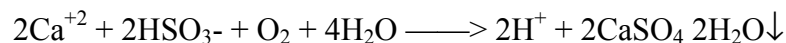
Small amounts of formic acid are added to buffer the slurry and keep it in a 4 to 5 pH range throughout the entire spray zone. Formic acid ionizes to form hydrogen and formate ions according to the following formula:



If the pH of the solution starts to decrease, then formate ions combine with H^+ , thus removing hydrogen ions and raising the pH. Alternatively, if the pH of the solution starts to increase, formic acid ionizes to provide hydrogen ions and lower the pH. Operating in the 4 to 5 pH range ensures the formation of easy-to-oxidize calcium bisulfite, the only water-soluble, calcium-sulfur salt (see Figure 4). Production of bisulfite avoids scaling and plugging in the absorber. Another function of formic acid is to react with limestone to increase the concentration of calcium ions in the recycle slurry, thus improving SO_2 absorption:



Air is bubbled through the absorber's reaction tank to oxidize the bisulfite ions to sulfate ions, which then react with calcium ions to form hydrated calcium sulfate (gypsum), which precipitates.



For all load conditions, the S-H-U process with its buffered slurry operates within the pH range that precludes sulfite formation. This greatly reduces the operating and maintenance requirements compared to unbuffered processes. An unbuffered process can require frequent maintenance and suffer reduced availability because of forced outages to clean the absorber. The ability to operate

in the nonscaling mode, even during transients, may be the biggest advantage to low pH-buffered absorption.

The buffered operation of the cocurrent/countercurrent absorber (see Figure 5) permits the absorption/oxidation reaction to occur at a much lower pH than in unbuffered absorbers. The large gypsum crystals that form in the scrubber reaction tank are easy to dewater and wash (purify) and are attractive to wallboard manufacturers because they produce a high-quality wallboard.

At Milliken, flue gas from the ESPs passes through new ID fans to the top of the cocurrent section of the absorber, where it is cooled and saturated by water sprays. The gas flows downward as it is contacted with recycle limestone slurry from spray nozzles at four levels (three plus a spare) to absorb SO₂. At the bottom of the cocurrent section, the recycle slurry disengages from the flue gas and collects in the absorber reaction tank (see Figure 3). The flow direction of the flue gas is reversed by rubber-lined turning vanes, and the gas flows upward through the countercurrent section, where it is contacted with recycle limestone slurry from spray nozzles at three levels (two plus a spare) to complete SO₂ absorption. The flue gas then passes through a two-stage mist eliminator before being discharged to the wet stack, which sits directly above the absorber. Wash-water sprays clean the upstream and downstream faces of both mist eliminators. The Milliken project provided the first demonstration of the S-H-U process installed directly beneath a plant stack, a design approach that saves considerable plot area.

Limestone is the primary reagent for reaction with absorbed SO₂. The equipment for limestone grinding, fresh limestone slurry storage, and fresh slurry transfer is located in the limestone preparation area of the FGD building. Limestone slurry is continuously pumped from the fresh slurry feed tank to the absorbers.

II.D.2 Gypsum Production

The recycle slurry from both cocurrent and countercurrent absorber sections collects in the absorber reaction tank. Although some oxidation occurs in the absorber as a result of excess oxygen in the flue gas, the reaction tank acts as a back-mixed reactor to complete oxidation of calcium bisulfite to calcium sulfate (gypsum). Blowers inject air, and side-mounted agitators provide complete mixing of air and slurry and prevent gypsum particles from settling to the bottom of the tank. The slurry in the reaction tank contains approximately 10-percent gypsum solids, which act as seed crystals for the formation of gypsum particles. This eliminates uncontrolled gypsum growth on absorber internals (gypsum scaling).

A slurry slipstream is pumped from the absorber sump to the dewatering area, which contains primary and secondary hydrocyclones, centrifuges, and other equipment. The slurry is concentrated to approximately 45-percent solids in the hydrocyclones and further dewatered to greater than 90-percent solids in the centrifuges. Fresh water is used to wash dissolved solids from the cake before it is discharged from the centrifuges. The gypsum cake is transported to a storage building, where it is stored until loaded onto trucks for delivery to the purchaser.

II.D.3 Blowdown Treatment

Essentially all the chloride in the flue gas is removed in the S-H-U scrubber; therefore, chloride must be removed to prevent it from building up in the slurry. A slipstream from the secondary hydrocyclone overflow is processed to control the concentration of chloride in the absorber slurry. The slipstream is first chemically treated to promote coagulation, flocculation, and sedimentation of suspended solids and dissolved metals. Reject solids are dewatered by a plate and frame filter press and disposed of in an approved permitted landfill. The chemically treated and clarified water can be discharged or further treated by a brine-concentration system in which 90 percent of the water is recovered as distilled water that is then used as makeup water for the FGD system. The remaining solution is a highly concentrated brine containing mainly calcium chloride, but also some magnesium and sodium chlorides, which can be sold commercially.

II.D.4 Stebbins Tile Absorber

An additional feature of this demonstration project is the use of a tile-lined, split-module absorber provided by the Stebbins Engineering and Manufacturing Company. The absorber is a concrete vessel with tile lining and a center dividing wall that divides it in half, thus providing each of Milliken's units with its own absorber module. Because each side of the vessel operates independently of the other, this innovative concept provides greater operating flexibility and reliability. The split module allows the flue gas from each boiler to be independently treated at a lower capital cost than would be required for two separate vessels. Tile lining provides superior abrasion and corrosion resistance when compared to rubber and alloy linings and should last for the life of the plant.

II.D.5 Heat Pipe Air Preheater System

An important feature of this project is the demonstration of the energy savings provided by a heat pipe air preheater, which was installed to replace the Ljungstrom regenerative type air preheater on Unit 2. The heat pipe design has the potential to eliminate many of the problems associated with tubular and Ljungstrom air preheater designs and to operate at a somewhat lower flue gas outlet temperature to improve overall plant heat rate. Heat pipe air preheaters operate as regenerative exchangers in which heat from the hot flue gas is indirectly transferred to the cold air by means of a working fluid that partially fills the heat pipe tubes. Passing hot flue gas over the lower portion of the tube causes the working fluid to boil. The vapor flows to the upper portion of the tube, where cold air flowing over the tube condenses the vapor, releasing latent heat, which warms the air. The heat pipes are mounted at a slight angle from horizontal, so the condensed liquid flows by gravity back to the evaporator end of the pipe to repeat the cycle.

A full-scale heat pipe air preheater consists basically of two ducts with a common wall (see Figure 6a). Individual heat pipe tubes (see Figure 6b) extend through the common wall across both ducts. Hot flue gas flows through one duct, while cold combustion air flows through the other duct. The tubes are sealed at the wall to prevent leakage from one duct to the other, thus improving efficiency. Fins are usually added to increase the surface area and heat transfer of the individual tubes.

II.D.6 Combustion Modifications

LNCFS™ III, a proven and reliable low-NO_x system furnished by ABB Combustion Engineering, was installed on both Milliken units 1 and 2. This was the first retrofit application on a utility boiler producing salable fly ash and equipped with a FGD system producing by-product gypsum. The carbon level in the fly ash is critical to fly ash sales and could adversely affect gypsum color. Therefore, it was crucial that the combustion modifications did not increase unburned carbon in fly ash beyond 4 percent.

LNCFS™ III uses a combination of bulk furnace staging and early, controlled coal devolatilization. Bulk furnace staging retards air and fuel mixing by diverting a portion of the combustion air, which is introduced at the fuel-burning zone. With conventional tangential firing, the introduction of excess combustion air during the early stages of coal devolatilization contributes significantly to the formation of NO_x. LNCFS™ III maximizes the bulk staging concept by using both overfire air and concentric firing. Staged combustion is accomplished by introducing a portion of the secondary air above the primary firing zone as overfire air. The concentric firing system utilizes a redirection of the secondary air, diverting it away from the coal stream. This reduces the combustion stoichiometry by preventing the fuel from entraining with the air during the initial stages of combustion. Fuel nitrogen conversion is reduced while maintaining appropriate oxidizing conditions along the furnace walls. Later, sufficient air is added to complete combustion and reduce carbon in ash.

Another important design feature incorporated into LNCFS™ III is the technique of early fuel ignition. Initiating the combustion point very close to the fuel nozzle produces a stable flame, which is more easily controlled under substoichiometric firing conditions and also contributes to lower thermal NO_x production.

II.D.7 Plant Economic Optimization Advisor

PEOA™, an on-line performance support system developed by DHR Technologies, Inc., was installed on both units 1 and 2. The system integrates key aspects of plant information management and analysis to assist plant personnel in meeting the requirements of the 1990 Clean Air Act Amendments (CAAA). The system optimizes the overall plant economic performance, including performance of the steam generator and turbine equipment, emissions systems, heat-transfer systems, auxiliary systems, and waste-management systems.

The PEOA™ automatically determines and displays key operational and control set points to minimize operating costs. The system provides operators with on-line emissions monitoring and diagnostic capabilities, along with rapid access to reports and trend information. The PEOA™ optimization algorithms evaluate key data emissions parameters, such as NO_x, SO₂, O₂, CO, and carbon dioxide (CO₂) levels, carbon in ash, and opacity, plus other operational parameters. PEOA™ also provides parameters or set-points to aid in the operation and performance of the unit.

II.E Project Implementation

The centerpiece of the demonstration facility is the S-H-U FGD plant, located in a new building about 100 ft from Unit 2. This building houses most of the equipment associated with the FGD process. It includes the absorbers, recycle pumps, oxidation air blowers, gypsum dewatering hydrocyclones and centrifuges, limestone storage silos, wet grinding equipment, blowdown pretreatment and brine concentration equipment, process water makeup tank and pumps, control room, electrical equipment rooms, and process laboratory. The building is 184 ft tall, and the roof-supported steel stack shell rises another 141 ft. Three stack flues extend another 50 ft, for a total height of 375 ft. Elevated ducts carry flue gas to the S-H-U plant from the Unit 1 and Unit 2 replacement fans, which are situated just downstream of their respective electrostatic precipitators (ESP).

II.E.1 Limestone Handling and

The limestone-preparation system is shown in Figure 7. Limestone is delivered to the site by truck and stored in an outdoor pile that has a three-month capacity. Front-end loaders reclaim the limestone and dump it into a reclaim hopper, where it is screened to minus six inches before being fed onto an enclosed conveyor for transfer to the storage bins in the FGD building. From the storage bins, the limestone flows to the two-train limestone-preparation system, each train with a capacity of 24 tons/hr of limestone slurry. One train, operating 12 hr/day, can supply the needs of both absorbers running at design rate.

At the heart of each train is a Fuller Traylor ball mill, consisting of a horizontal steel cylinder with a replaceable rubber liner. Each ball mill is charged with 75 tons of steel balls. Limestone, metered into each mill by a gravimetric feeder, is ground with clarified water, and the limestone slurry is discharged through the mill's trommel screen into an agitated rubber-lined product tank, to which clarified water is added to adjust slurry density. The slurry flows to the classifier, which has five small hydrocyclones and three large hydrocyclones. The small hydrocyclones are used to produce slurry with a fineness of 90 percent through 325 mesh, while the larger hydrocyclones produce a fineness of 90 percent through 170 mesh, the size required when formic acid is being used in the FGD scrubbers. The classifier overflow can be sent to either the Unit 1 or Unit 2 feed tank, and the underflow is returned to the ball mill. From the feed tanks, the slurry is pumped to the absorbers at a flow rate that depends on the SO₂ loading in the flue gas.

II.E.2 Formic-Acid-Handling System

The major components of the formic-acid-addition system are the delivery truck unloading station, the storage tank, and the metering pumps. Formic acid is added as needed to maintain the pH in the absorber in the desired range.

II.E.3 Absorber

The concrete 42 ft by 79 ft by 108 ft high absorber vessel is the heart of the S-H-U FGD plant. An internal partition from the floor to the top divides this vessel into two identical modules, each with independent flue gas inlet and outlet connections. Each of these modules has an internal

partition that extends from a point 45 ft above the floor to the top and divides the module into cocurrent and countercurrent sections. The bottom 20 ft of each module forms the absorber reaction tank. The vessel walls and floor are completely lined with Stebbins ceramic tile. To protect the tile from thermal shock, the top 28 ft of the walls in the cocurrent section are also lined with acid and shock-resistant brick.

A dedicated recycle pump pumps slurry to each absorber spray bank directly from the absorber reaction tank. Each level of spray nozzles has its own pump, and individual pumps can be taken out of service to handle reduced flue gas flow rates. Seven pumps are provided for each module, three operating and one spare on the cocurrent side and two operating and one spare on the countercurrent side. Each absorber reaction tank has five agitators to maintain solids in suspension and to mix oxidation air with the absorber slurry to force the bisulfite oxidation reaction. Oxidation air is injected into the slurry through air lances mounted just in front of four of the five agitators. The system is designed to remove 95 percent of the SO₂ when burning 3.2-percent sulfur coal with five of the seven recycle pumps operating. Operation of six of the seven pumps, plus the use of formic acid is necessary to achieve 98-percent SO₂ removal.

II.E.4 Induced Draft (ID) Fans and Stack

New induced draft fans were installed to provide the higher pressure needed to overcome the resistance of the S-H-U FGD system. Each unit has two 50-percent capacity fans. Because of the installation of the essentially leak-free heat pipe air preheater on Unit 2, the ID fans for Unit 2 are slightly smaller than those for Unit 1. Dedicated ducts deliver flue gas from the ID fans to the FGD plant. However, the Unit 1 and Unit 2 ducts are interconnected by a crossover duct, which permits, at reduced load, flue gas from both boilers to be treated by either absorber module.

Each absorber module discharges treated, saturated flue gas to the atmosphere through the stack. The stack has three independent flues, one 12-ft diameter fiberglass reinforced plastic (FRP) wet flue dedicated to each absorber module and a common 8-ft diameter steel bypass flue. The three flues are supported inside a 40-ft diameter steel stack that is mounted on the roof of the FGD building.

II.E.5 Gypsum Dewatering

Gypsum solids that form in the absorber reaction tanks are purged from the system through the gypsum dewatering system, which produces gypsum cake suitable for use in the manufacture of wallboard (see Figure 8). A small fraction of the recovered liquor is processed by the FGD blowdown treatment system to control chloride accumulation; the rest is recycled. Each absorber has a bleed pump to feed absorber slurry to the primary dewatering hydrocyclones. The concentrated gypsum slurry from the primary hydrocyclones is processed by centrifuges to produce a gypsum product meeting specifications for purity and surface moisture. Centrifuge filtrate passes through the basket, exits the casing, and is piped to the filtrate tanks. After the gypsum has been washed and spun dry, the basket speed is reduced and the scraper is gradually rotated into the cake. As the cake is peeled from the inside surface of the basket, it falls through the discharge chute onto the discharge conveyor. The gypsum-handling system transfers gypsum

cake containing 6- to 8-percent surface moisture from the centrifuges to the gypsum storage building. A front-end loader loads the gypsum onto trucks for shipment to the end user.

The overflow from the primary dewatering hydrocyclones is collected in the secondary hydrocyclones feed tanks. Part of this slurry is processed by the secondary hydrocyclones to remove most of the residual gypsum to produce clarified water for use in limestone slurry preparation, FGD system flushing, and further treatment for chloride control. The balance is recycled to the absorbers.

II.E.6 FGD Blowdown Treatment

Chlorides in the coal, which end up in the flue gas, are scrubbed out by the FGD system. Because commercial grade gypsum specifications limit the allowable chloride content, the concentration of chlorides in the scrubbing liquor (the design level is 40,000 ppm) must be controlled. This is accomplished by treating a slipstream from the clarified water tank. The blowdown pretreatment system (see Figure 9) removes suspended solids and heavy metals from the blowdown prior to discharge or processing for brine concentration. Removal of these constituents is necessary to meet discharge permit requirements and to maintain salable quality calcium chloride as a brine product.

Three reaction tanks are connected in series. Lime slurry is added to the first tank to neutralize the mineral acids present and initiate metal hydroxide precipitation. The second tank is the heavy metal precipitation tank where an organosulfide can be added, if needed, to achieve further heavy metal precipitation. The third tank is the coagulation tank, where a mineral coagulant, ferric chloride, is added to the blowdown stream. From the coagulation tank, the blowdown stream flows to the Densadeg® unit, which has three stages: a solids contact reaction zone, a presettler/thickener zone, and a lamellar settling tube clarifier zone. The reaction zone is a rubber-lined tank with a high energy axial flow mixer. Anionic polymer flocculent is added, and wastewater is recirculated by the mixer, densifying the precipitated metal hydroxide particles. The wastewater is forced upward by the mixer, over a weir and through a spillway to the presettler/thickener zone.

The presettler/thickener zone and the clarifier zone are housed in a tall cone-bottomed tank. Sludge is withdrawn from the tank at two locations. Sludge is withdrawn from the center well and transferred to the sludge holding tank prior to dewatering by the filter press. After dropping out its precipitated solids, the blowdown stream flows to the brine concentrator storage tank. Waste sludge from the brine concentrator underflow pump is also stored in the tank. A plate and frame filter press is used to dewater the FGD blowdown treatment sludge.

II.E.7 Brine Concentration System

The FGD brine concentration system (see Figure 9) removes chlorides from the FGD blowdown pretreatment system effluent, producing a concentrated calcium chloride brine that can be commercially marketed, plus a pure-water stream that is recycled. Hydrochloric acid and scale inhibitors are added to the brine concentrator feed tank to prevent carbonate and sulfate scaling of the system. From the feed tank, the brine is pumped through the primary heat exchanger,

where the brine is preheated to 150 °F, and into the deaerator. From the deaerator, the brine is pumped through the secondary heat exchanger, where it is heated to 209 °F, to the brine concentrator, which is a vapor compression type falling film evaporator. The brine concentrator was designed to produce clean distilled water for recycle and a concentrated brine by-product for sale.

As water is evaporated from the recirculating brine solution, the concentration of the remaining brine increases, and salts start to precipitate. The first salt to precipitate is calcium sulfate; calcium sulfate crystals are removed by a hydrocyclone and recycled. If the brine becomes too concentrated, other salts precipitate, but this is prevented by controlling the salt concentration in the by-product brine.

II.E.8 Heat Pipe Air Preheater

The two vertical flow Ljungstrom air preheaters originally on Unit 2 were replaced with two vertical flow heat pipe preheaters provided by ABB Air Preheater. Each heat pipe unit consists of 12 heat-transfer modules. To prevent leakage, tubes in the modules are seal welded to the dividing wall between the air and flue gas sections. The tubes are tilted 5° from the horizontal. Because the tubes are fixed only at the dividing wall, they are free to expand and contract without causing a problem. Soot blowers are provided to keep the pipes clean. Although there was no leakage between the flue gas and air sides of the heat pipe, the soot blowers were loose-fitted, and some gas leakage occurred around their penetrations.

II.E.9 ESP Upgrade

As part of the MCCTD Project, the ESPs on units 1 and 2 were upgraded to accommodate the wet FGD system. Originally, the particulate-control system consisted of two ESPs in series, stacked one on top of the other. Each ESP consisted of two independent sections with the gas flow separating upstream of the air preheater and rejoining downstream of the final ESP. During the modifications, the bottom ESP was removed completely, and the top ESP was rebuilt. The internals were replaced using a wide spacing design provided by Belco Technologies, Inc. With 16-inch plate spacing, the modified unit is smaller and requires less energizing power.

II.E.10 Makeup Water

Makeup water is required for the mist eliminators, absorber wall wash to prevent buildup at the wet/dry interface, quenching to cool the gas prior to absorption, and for washing the by-product gypsum cake. Makeup water is also added to the absorber reaction tank to maintain a water balance. The makeup water comes from Milliken's Process Water Reclamation Facility, with the power plant's service water system serving as backup.

II.E.11 System Control

The rate of limestone addition to the absorber reaction tank is controlled by a feed-forward control system. The flue gas flow rate is multiplied by the SO₂ concentration to generate a signal proportional to the total SO₂ flow to the absorber, and limestone slurry is added to the absorber in

direct proportion to this value. The pH of the slurry is monitored to ensure that the limestone control system is working correctly. Formic acid is added to the absorber reaction tanks using variable capacity piston pumps. Pump rate is set manually based on laboratory analyses to maintain the desired formic acid concentration in the reaction tank. The recycle pumps have only on/off control; there is no control on throughput. These pumps are taken out of service or placed on line at the operator's discretion to suit load conditions.

Each absorber module is equipped with two mist-elimination stages, which are located above the last absorber recycle spray bank in the countercurrent section of the absorber. The mist elimination wash system provides online cleaning by spraying the mist eliminators with makeup water.

To oxidize bisulfite ions to sulfate, a constant amount of compressed air is injected into the absorber reaction tanks through air lances located adjacent to the absorber agitators. Water is injected into the air supply header to saturate and cool the air prior to injection in order to prevent formation of "wet/dry" buildup and consequent air lance plugging.

The gypsum bleed and dewatering system removes gypsum solids from the FGD system, maintaining the suspended solids concentration in the absorber reaction tank within a preset range. Density of the reaction tank slurry is sensed in the bleed pump discharge line. The centrifuges are controlled in accordance with a preprogrammed six-step batch cycle.

II.F Project Objectives and Statement of Work

The general objective of this project, as listed in the Cooperative Agreement awarded on October 20, 1992, was to demonstrate high-sulfur capture efficiency and NO_x and particulate control at minimum-power requirements, the production of by-products in lieu of wastes, and zero waste water discharge. The Statement of Work (SOW) states that "the Milliken Station is currently among the top 20 power plants in the nation for overall heat rate performance, and the project is designed to maintain that excellent rating. Innovative features to be demonstrated include S-H-U's advanced SO₂-scrubber technology, which uses formic acid enhancement and cocurrent/countercurrent open spray tower absorber design; tile-lined split module absorber construction, as provided by Stebbins; NO_xOUTTM injection and air combustion modeling technology and implementation for NO_x control; and heat pipe air preheater technology to increase energy efficiency and conservation." As stated previously, NO_xOUTTM was subsequently deleted from the demonstration.

The SOW covers the modifications to Milliken Station, including both upgrades and new facilities, and assigns various responsibilities. Specific facility modifications and additions listed in the SOW are:

- FGD System with design basis provided by S-H-U and structural basis for the absorber vessel provided by Stebbins
- Limestone-Handling Facilities

- FGD Blowdown Treatment Facilities including blowdown neutralization/precipitation and brine concentration.
- Gypsum Pelletizing and Storage Facilities
- Units 1 and 2 ESP Upgrades to improve performance
- Units 1 and 2 Upgrade for NO_x removal by adding new coal pulverizers and new low-NO_x burners and other equipment
- Unit 2 Urea Injection (later deleted from the project)
- Unit 2 Heat Pipe Air Preheaters to replace the Unit 2 Ljungstrom air preheaters
- Units 1 and 2 Station Service, ID Fans, and Ductwork Upgrade

The project was divided into three phases: Engineering/Design, Construction, and Operations. This report deals mainly with Operations. The SOW declares that NYSEG will oversee operations progress and management and will take all necessary steps to ensure that the Test Plan submitted to DOE is accomplished. However, the SOW does not cover details of the tests to be performed. These are covered in a test plan, which the SOW states, is to be developed (CONSOL, 1996). All activities related to data analysis and reporting, including completion and submittal of all government-required technical reports, are included in this task.

III Technical and Environmental Assessment

A major objective of the project is to determine the effect modifications have on the Milliken Station's performance, such as pollutant emissions and efficiency. Another objective is to determine the effect of various parameters on scrubber performance. Process variables whose effects were studied include coal sulfur content, formic acid concentration, spray header operating combinations, gas velocity, liquid-to-gas (L/G) ratio, and limestone grind size.

III.A Technical Results

III.A.1 LNCFS™ III Performance

Because there was not enough time to do baseline testing on Unit 1 prior to installation of the LNCFS™ III burners, it was decided to compare baseline performance of Unit 2 with post-conversion performance of Unit 1. Previous studies had indicated there was very little difference in NO_x emissions between the two units; therefore, the comparison should be valid. Performance evaluation is based on long-term (60-70 days) testing and includes the impact of the LNCFS™ III system on NO_x emissions, boiler efficiency, carbon on fly ash, and CO emissions. Specific conclusions were:

- At full boiler load (145-150 MWe) and 3.0- to 3.5-percent economizer O₂, the LNCFS™ III system lowered NO_x emissions from a baseline value of 0.64 to 0.39 lb/MBtu, which is a 39-percent reduction. At a reduced boiler load of 80-90 MWe and 4.3- to 5.0-percent economizer O₂, the LNCFS™ III system lowered NO_x emissions from a baseline value of 0.58 to 0.41 lb/MBtu, a 29-percent reduction. These results are within the expected range.
- Boiler efficiency was 89.3-89.6 percent for baseline and 88.3 to 88.5 percent for the LNCFS™ III system. The LNCFS™ III boiler efficiency was lower than baseline because of higher post-retrofit flue gas O₂ levels and higher stack temperatures, which accompanied the air preheater retrofit. When the LNCFS™ III system and baseline results were adjusted so that they could be compared at similar flue gas temperatures and compositions, the estimated LNCFS™ III boiler efficiency was 0.2-percent higher than baseline.
- With the LNCFS™ III system, carbon in fly ash below 4 percent was maintained, and CO emissions did not increase.

III.A.2 S-H-U Process

A series of tests were planned for the S-H-U scrubber to determine the effect of a number of variables on scrubber performance. Parameters to be varied included coal sulfur level, L/G ratio, gas velocity in the cocurrent zone, formic acid concentration, and limestone grind size. The main variable to be studied was the L/G ratio, which was altered by varying the number of spray

headers in operation. Each spray header is fed by its own pump, but these pumps are not of variable speed and only operate in on/off mode; there is no flow control on the headers. Therefore, a header is either in operation or it is not. As discussed earlier, there are four headers in the cocurrent section (A, B, C, and D, with A being the top header) and three headers in the countercurrent section (E, F, and G, with E being the top header). A test condition is indicated by two numbers in parentheses, where the first number is the number of cocurrent headers in operation, and the second number is the number of countercurrent headers in operation, for example, (4, 2) means four cocurrent and two countercurrent headers in operation.

Tests were planned for coals with three different sulfur levels (see Table 1), with the initially expected sulfur levels being:

- Low-sulfur coal (1.6-percent sulfur)
- Design-sulfur coal (2.9-percent sulfur)
- High-sulfur coal (4.3-percent sulfur)

Table 1. Properties of Design Coals From Public Design

Proximate Analysis, wt%	Medium Sulfur	Base Coal	High Sulfur
Moisture	6.3	5.6	7.5
Ash	10.9	9.5	10.5
Volatile Matter	29.0	37.1	37.4
Fixed Carbon	53.8	47.8	44.6
Higher Heating Value, Btu/lb	12,600	12,800	12,165
Sulfur, wt%	1.6	2.9	4.3
Grindability, HGI	73	57	56
Ash Fusion Temperature, °F	2480	2200	2222
Ultimate Analysis, wt% (dry)			
Ash	10.9	9.5	10.5
Carbon	73.5	74.3	66.4
Hydrogen	4.3	4.5	4.6
Nitrogen	1.3	1.5	1.2
Chlorine	0.1	0.2	0.1
Sulfur	1.6	2.9	4.3
Oxygen	2.0	1.5	5.4

¹ Properties of the coals that actually burned during the MCCTD Project differed somewhat from those used as the design basis.

The coals actually burned had somewhat different sulfur levels (1.6 percent, 2.23 percent, and 3.2 to 4.1 percent, respectively). Originally, an extensive series of tests was planned, but for various reasons, many of these tests were not performed. Only a summary of results is presented in the following sections, and the figures are intended only to give typical results. For a more detailed analysis of the results, consult the appropriate reports (CONSOL, 1999a; New York State Electric & Gas Corporation, 1999; CONSOL, 1999b; CONSOL, 2000) **Results With Low-Sulfur Coal**. This test was run on Unit 2 from October 11 to November 21, 1995. The coal burned during this period had a sulfur content of 1.6 percent. Gas velocity varied from 12 to 20 ft/s. Results are shown in Figures 11-15. Conclusions reached were:

- SO₂ removal ranged from 30 percent, using only two spray headers without formic acid addition, to 98 percent, using all seven spray headers and formic acid at a concentration of 800 ppm (see Figure 11).
- Adding formic acid significantly enhanced SO₂ removal. With five spray headers in operation, SO₂ removal increased from 82 percent without formic acid to 97 percent with 800 ppm of formic acid.
- The pH of the spray slurry affected the results. For example, with all seven spray headers in operation, SO₂ removal was 90.2 percent at a pH of 4.21 and increased to 94.4 percent at a pH of 5.05.
- Although there was some variation depending on operating conditions, on average, SO₂ removal increased 2.6 percent when a finer limestone grind size (90 percent through 325 mesh compared to 90 percent through 170 mesh) was used (see Figure 15).
- At equivalent L/G ratios, the unit performed better at high gas velocity than at design velocity. For example, SO₂ removal averaged 95 percent at 94 gal/1000 acf at the design velocity and 97 percent at 89 gal/1000 acf in the high-velocity tests. That is, at a given L/G, higher mass-transfer rates were achieved at gas velocities above the design value.
- Changing the L/G ratio in the cocurrent section had no measurable effect on pressure drop, but increasing the L/G ratio in the countercurrent section significantly increased the absorber pressure drop.
- Mass transfer (and, consequently, SO₂-removal rate) increased with increasing L/G, but the effect was not necessarily linear (see Figure 12).
- Formic acid increased the mass transfer rate, but the effect was not linear and diminished with increasing formic acid concentration (see Figure 13).
- There is little effect of chloride level on SO₂-removal efficiency as long as chloride concentration is below about 5 percent (50,000 ppm) (see Figure 14).

Results with Design Coal. Tests of the S-H-U FGD process, while burning design coal in Unit 1 at full load, were conducted November 12-20, 1998. These tests were originally scheduled to last

for 100 days and vary the following parameters: L/G ratio, gas velocity, formic acid level, and limestone grind size. However, problems with the spray nozzles (plugging and breakage) delayed this test series. Also, there were problems obtaining a supply of coal with the design sulfur level. Finally, in order to complete the tests during 1998, the test plan was considerably truncated. Deregulation by the State of New York led to the sale of the Milliken plant, and once the pending sale was announced, there was considerable pressure to complete the test program. The revised test plan eliminated gas velocity, formic acid level, and limestone grind size as variables. Thus, the only variable tested was L/G ratio, which was varied by changing the number of spray headers in operation. The average sulfur content of the coal burned during these tests was 2.23 wt%.

The major conclusions from this test period are:

- L/G ratio is an important variable with SO₂ removal ranging from 85.6 percent with only five spray headers in operation to 95.1 percent with all seven headers in operation. These results are with a formic acid concentration of nominally 800 ppm, a limestone grind size of 90 percent through 170 mesh, and a gas velocity of 20 ft/s in the cocurrent zone.
- Although the pH of the system was maintained at 4.1 ± 0.1 for most tests, one test was run at a pH of 3.9. For this test, SO₂ removal was 85.4 percent compared to 91.5 percent for a test at the same operating conditions except for a pH of 4.1. This is an indication that pH is a significant variable and that maintaining pH in the specified range is important for achieving maximum SO₂ removal.
- Pressure drop across the scrubber was a function of L/G ratio in the countercurrent zone, but almost independent of L/G in the cocurrent zone.
- For the tests in which not all seven spray headers were operating, SO₂ removal depended on which spray headers were not operating. Results with six of the seven headers in operation are shown in Table 2 (cocurrent headers are A, B, C, D, and E and countercurrent headers are E, F, and G, with A and E being the top headers in their respective zones):

Table 2. Spray Header Results

Header Not in Operation	SO₂ Removal, %
A	86.7
B	93.4
C	90.7
D	91.5
E	87.3
F	90.2
G	91.7

From these results, it appears that the top header in each zone is the most important. This is logical, since the spray from the top nozzles would have the longest contact time with the flue gas.

Results with High-Sulfur Coal. The test of the S-H-U process when burning high-sulfur coal was performed on Unit No. 1 between April 28 and May 1, 1998. Sulfur level of the coal burned during this period ranged from 3.2 to 4.1 wt% and averaged 3.64 wt%. The original plan for high-sulfur coal testing included eight tests over a four-day period. Only this short period of operation with high-sulfur coal was scheduled to reduce the risk of a long period of operation with high SO₂ emissions. Because of difficulty controlling pH during these tests, the original test plan was not adhered to, and some of the planned tests were eliminated.

During the high-sulfur coal test, most operating variables were kept constant: Unit 1 operated at full load, chloride level in the scrubber slurry was about 4 wt% (40,000 ppm), the limestone grind size was 90-percent less than 170 mesh, and the gas velocity in the cocurrent section was 20 ft/s. Unit No. 2 did not operate during the high-sulfur test. The pH-control loop was set to maintain a pH of 4.8, but control was difficult because a high limestone-slurry-feed rate, beyond pump capacity, was required, and because scrubber chemistry was affected by a large amount of fly ash that escaped the ESP and ending up in the scrubber.

The main variable to be studied in this test was the effect of liquid-to-gas ratio on sulfur-removal efficiency. Three combinations of headers were used in this test: (4, 3), (3, 3), and (3, 2). A planned (2, 3) test, as well as replicates of the other tests, was deleted because of operating problems. After the three tests listed above, for the rest of the time on high-sulfur coal, the absorber ran in the (4, 3) configuration.

Conclusions reached in this test are:

- Sulfur removal varied widely, ranging from 43.7 percent at a pH of 3.15 to 97.6 percent at a pH of 4.85. Thus, the pH of the limestone slurry appears to have a significant effect on sulfur removal. In fact, SO₂ removal depended more on pH than on scrubber L/G ratio. During this test period, pH was out of the control range much of the time because of low limestone slurry flow caused by inadequate pumping capacity for the limestone demand with the high-sulfur coal.
- As expected, sulfur-removal efficiency was lower than for the low-sulfur coal and design-sulfur coal tests.
- The pH control problem was mainly attributed to the inability of the limestone slurry feed system to keep up with demand. This problem arose because the system was not designed to handle 4-percent sulfur coal. The fact that SO₂ removal was as high as 97.6 percent when the pH was 4.85 suggests that the S-H-U process can handle higher-sulfur coal if designed to do so.

III.A.3 ESP Performance

CONSOL conducted performance tests on the original and modified ESPs (CONSOL, 1996). The same coal was fired during these tests. Results indicate that the modified ESP shows better removal efficiency than the original unit, even though it has less than one-half of the collection plate area of the original ESP. The data indicate that the power requirement is 25 percent less than that of the original ESP.

III.B Environmental Performance

The Milliken S-H-U project, through installation of a variety of innovative emissions-control technologies, was designed to comply with all applicable Federal and New York State Department of Environmental Conservation air, water, and solid-waste environmental regulations. In most cases, pollutant levels were reduced when the MCCTD Project was operated. No problem areas were identified concerning environmental regulations or permit conditions caused by the operation and performance of the equipment being demonstrated under the MCCTD Project.

III.B.1 Air Emissions

Air emissions controls for the project consisted of a high-efficiency FGD system for SO₂ control, LNB's for NO_x control, and ESP and coal mill upgrades for particulate control. New continuous emissions monitoring systems were installed to ensure continuous measurement of stack emissions, and stack testing for SO₂, NO_x, total suspended particulates, CO, and opacity was performed for each boiler in accordance with NYSDEC permit conditions. NYSEG also installed a monitoring system to collect ambient air quality data for SO₂, NO_x, and particulates.

SO₂ and NO_x emissions were reduced by operation of the MCCTD Project, and the resultant acid-rain reduction had a positive impact on acid-rain-sensitive areas such as the Adirondack and Catskill mountains. The project significantly reduced measured concentrations of SO₂ at the NYSEG ambient air monitoring stations (57 to 72 percent, depending on monitor location). NO_x and O₃ concentrations were moderately reduced at the majority of sampling sites. Although average particulate matter emissions concentrations were well below the NYSDEC permit limit, no conclusive changes in ambient levels for particulates were detected. Measured levels of SO₂, NO_x, O₃, and particulates were below National Ambient Air Quality Standards throughout the entire ambient-air-monitoring program.

Emissions of CO₂ increased slightly (less than 2 percent) as a result of the project because of additional power consumption by the scrubber and because CO₂ is released when limestone reacts with SO₂.

III.B.2 Water Discharges

One of the project's goals was to achieve zero-wastewater discharge. Major station elements that generate wastewater include cooling water systems, boiler blowdown, demineralizer backwashes, sump pump discharges, and sanitary sewage. The remainder of the wastewater is comprised of regeneration wastes, water from washing paved areas, yard and roof drainage, and drainage from the coal storage pile and ash landfill. Sanitary waste is discharged separately. Problems with the brine-concentration system necessitated temporary approval from the NYSDEC to discharge the brine stream into Milliken's Process Waste Reclamation Facility.

All water discharges to Cayuga Lake complied with the existing State Pollution Discharge Elimination System Permit. No significant impacts to potable water, surface water, or ground water resulted from the project. No new intake/discharge structures were required, and thermal discharges to Cayuga Lake remained unchanged. Because the MCCTD Project did not affect any thermal discharges to Cayuga Lake, aquatic ecological resources associated with Cayuga Lake remained unchanged.

III.B.3 Solid Waste

The solid wastes that resulted from the project were comprised of a small portion of the gypsum that was occasionally off-spec and, therefore, unmarketable, unmarketable chloride salts resulting from brine concentrator problems that were never fully resolved, limestone area runoff sediment, and FGD-blowdown treatment sludge. Handling, transportation, and disposal of these wastes met all current NYSDEC landfill permit requirements. NYSEG also operated a landfill groundwater-monitoring system to ensure that groundwater was not affected by the project. Monitoring wells down gradient of the operational, post-1984 section of the landfill indicated no groundwater contamination problems.

III.B.4 Waste Minimization/Recycle

The project had a strong waste minimization/recycle theme with a goal to produce marketable commercial-grade gypsum and calcium chloride brine by-products to minimize solid-waste disposal. Gypsum production proved to be highly successful, but the brine concentration system never became fully operational (see Section III.C.6). The new LNCFSTM III burners, once tuned, significantly reduced NO_x emissions while minimizing the amount of unburned carbon contained in the fly ash, preserving salability of the ash.

III.B.5 Air Toxics

An Air Toxics and Emissions Characterization Program was conducted after the MCCTD Project was installed. Removal efficiencies and system mass balances were determined for key air toxics compounds. The findings for this program are documented in a report prepared by Carnot (Janati, K., 1997).

III.C Impacts on Equipment

Both Units 1 and 2 started up in 1995 without problems and achieved the design 95-percent SO₂ removal efficiency within a few hours while burning 1.8- to 2.2-percent sulfur coal. The units met all the process guarantees, and the by-product gypsum has been of uniform quality and has been sold for commercial applications. Although performance has been generally satisfactory in over 30,000 hr of operation, there have been some problems that are discussed below.

III.C.1 Scaling and Plugging

The absorbers experienced a greater level of deposits on the absorber internals, slurry piping, and dewatering equipment than expected. Problems also occurred with spray nozzle plugging and plugging of the individual hydrocyclone elements of the primary dewatering system. The spray nozzles were made of ceramic material that was very erosion resistant but brittle; thus, they were prone to breakage during cleaning.

A number of causes were identified for these problems. Part of the plugging was caused by loss of rubber from the rubber-clad turning vanes in the absorbers. These rubber particles were picked up by the recycle and bleed pumps, which had no suction screens, with the result that downstream equipment (nozzles and hydrocyclones) became plugged. These problems can be significantly mitigated by redesign of the turning vanes and adding suction screens to the pumps.

Some of the equipment scaling was attributed to operating for extended periods at lower than design gypsum solids concentrations in the slurry. This lowered the inventory of seed crystals in the slurry and resulted in gypsum precipitating on equipment surfaces rather than on the seed crystals. This can be overcome by making sure that the gypsum concentration in the recycle slurry is at the proper level. Operation at higher solids level in the absorber not only reduced the amount of plugging in the spray nozzles and lessened scaling, but also greatly improved the operability of the hydrocyclones and centrifuges.

There is also evidence that scale formation was worse when a different limestone was used. Dewatering the gypsum when this limestone was used was also more difficult. This indicates that limestone properties are important and suggests that specifications for the limestone used with the S-H-U process need to be developed.

III.C.2 Settling in the Absorber Reaction Tank

Problems were encountered with solids settling to the bottom of the reaction tank and blocking the suctions of some of the pumps (see Figure 10). This required that these pumps be run continuously to keep their suctions open. This could result in lower efficiency (increased power consumption) for reduced load conditions, when some of these pumps might not need to be operating. The problem with settling was traced to the fact that with the split-module design, all the mixers for each module had to be mounted through one wall, since no mixers could be mounted on the center dividing wall. This resulted in inadequate mixing to keep the gypsum particles in suspension everywhere in the reaction tank, with the result that settling occurred at some locations. In future split module designs, mixer placement needs to be better engineered.

III.C.3 Spray Nozzle Breakage

As discussed above, the ceramic spray nozzles were difficult to clean without breakage. To overcome this problem, some of the nozzles were replaced with redesigned nozzles made of stellite. These nozzles worked well. However, it was found that after suction screens were installed on the recycle pumps, plugging of the spray nozzles was greatly reduced. Therefore, in future units, with suction screens installed on the pumps, it may be satisfactory to continue using ceramic nozzles.

III.C.4 Reagent Underutilization

Reagent underutilization was experienced under some operating conditions. The problem was traced to the fact that, in the Milliken design, both the reagent addition port and the gypsum bleed pump suction are located on the same wall. Because of the inadequate mixing problem discussed above, the proximity of the limestone-addition and bleed-withdrawal points meant that some limestone was being withdrawn before it had time to react. This problem was largely remedied by extending the limestone addition point further into the reaction tank. In future designs, the limestone addition and bleed withdrawal points should be as widely separated as possible.

III.C.5 Acid Brick Deterioration

Some deterioration of the acid brick in the absorbers in the region of the A and B headers has occurred. The cause of this deterioration has not been identified.

III.C.6 Brine-Concentration System

The brine-concentration system experienced numerous operating problems and did not work satisfactorily at any time during the demonstration. Problems encountered included excessive vibration in the vapor compressor, a high-boron concentration in the feed to the brine concentrator, suspended solids in the concentrated brine, scaling and plugging of the evaporator tubes, and corrosion in various parts of the system. After some process modifications, including changes to the vapor compressor, addition of a filter on the product brine, addition of sodium sulfate and sodium hydroxide to control scaling, and replacing some stainless steel parts with hastelloy, a 10-day trial run was performed in December 1997. The unit ran satisfactorily except that the impurities levels, such as boron, in the product brine were higher than allowed by product specifications. Thus, satisfactory operation of the brine concentration system was not achieved during the time frame of this project. However, if the impurities levels can be reduced, acceptable operation of the system should be achieved.

III.C.7 Stebbins Tile

Inspection of the absorbers after two years of operation found the tiles to be in good condition with no obvious erosion or deterioration. The tile grouting was hard and appeared to be undamaged in any way except for two locations, where short lengths of grout had eroded, apparently because of spray impingement. The three cracked tiles, originally purposely installed

on one of the walls, were replaced after a year's operation, with no leaks where the cracked tiles had been. A method for sealing leaks was tested and proved to be satisfactory.

III.C.8 Heat Pipe Air Preheater

Because of the heat pipe air preheater's design, there is no gas leakage between the flue gas and air sides. However, because the flue gas sides operate at a negative pressure, significant amounts of air can be drawn into the heat pipes through check valves on the soot blowers and at the soot blower wall penetrations. Tests indicate the air in leakage rate as 1.5 to 3 percent of the inlet flue gas flow. Heat pipe pressure drops met or exceeded design values. Substandard performance of the heat pipe air preheaters was initially experienced because of contaminated working fluid. After the working fluid was replaced with pure material, performance was satisfactory.

III.D Evaluation of Project Achievements

An overall assessment of the S-H-U process is that it performed well. Sulfur-dioxide removal levels as high as 98 percent were achieved under some operating conditions, and the unit operated routinely in the 90- to 95-percent SO₂-removal range. Although the test program was not as extensive as originally planned, tests on operating variables showed which parameters are most important in maintaining high SO₂ removals. Unit reliability was high, with availability near 100 percent. The compact design proved generally successful, although there were minor problems with locations of mixers, pump suction, and limestone inlets. Thus, a future unit would probably be designed somewhat differently. The gypsum produced was of high-quality, and essentially all of it was sold for wallboard manufacture. The brine-concentration system never worked satisfactorily, so that the goal of zero wastes was not achieved.

IV Market Analysis

The S-H-U process has been used in Europe to reduce SO₂ emissions from utility and industrial boilers firing coal, lignite, oil, and gas, as well as municipal waste incinerators and has wide applicability for SO₂ capture from flue gas within the U.S. utility and industrial sectors. The process is applicable to boilers firing low-, medium-, and high-sulfur coals regardless of boiler size or type and can achieve up to 98-percent SO₂ removal, provided all factors are taken into account during the initial design. All wet limestone FGD processes require significant plot areas, but those required by S-H-U can be minimized by adopting the split-module, below-the-stack configuration demonstrated in the MCCTD Project.

IV.A Market Size And Penetration

A detailed market analysis has been developed by NYSEG (New York State Electric & Gas Corporation, 1993). In this analysis, the total U.S. electric market for which the S-H-U process was applicable was divided into two parts: retrofit capacity (pre-NSPS coal-fired boilers not equipped with SO₂ controls) and new capacity (projected coal-fired additions through 2030). This analysis indicates a total potential retrofit market of 5,700 MWe through 2030 and a potential new power plant market of 96,200 MWe. Although done in 1995, this analysis should still be reasonably valid. The successful demonstration of the S-H-U process at Milliken, along with the experience in Europe, should enable S-H-U to effectively market its FGD technology in the United States through its U.S. design and manufacturing partners.

IV.B Economics

Because of differences in performance and site-specific characteristics, it is difficult to compare economics of the S-H-U process with other FGD processes. Nevertheless, some comparisons can be made. A study of 24 different FGD processes by the Electric Power Research Institute (EPRI) (Electric Power Research Institute, 1991) indicate that regenerable processes typically have higher capital and operating costs than limestone wet scrubbing processes. Advantages of the S-H-U process compared to regenerable processes include lower energy consumption, lower capital and annual operating costs, and minimal solid wastes and scrubber blowdown. However, compared to other limestone wet scrubbing processes, the S-H-U process appears to have higher capital costs.

IV.B.1 Capital Cost

The estimated capital cost for retrofit of the S-H-U process on a 300-MWe power plant burning 3.2-percent sulfur coal and operating at a 95-percent SO₂-removal efficiency is shown in Table 3.

Table 3. S-H-U Retrofit Capital Cost

	<u>\$10⁶</u>	<u>\$/kW</u>
Limestone Handling and Preparation	5.3	17.7
Slurry Feed and Recycle	3.7	12.2
Absorber Module and Auxiliaries	5.6	18.5
ID Fan and Ductwork	6.3	21.1
Stack and Flues	2.5	8.3
Gypsum Dewatering and Handling	4.2	14.1
Waste Water Processing System	2.4	7.9
Other Mechanical Systems	4.7	15.8
Electrical and I&C	5.6	18.7
FGD Building and Site Work	13.4	44.7
Total Process Capital	53.7	179.0
Engineering and Home Office (10% TPC)	5.4	17.9
Project Contingency (10% TPC + E&HO)	5.9	19.7
Total Plant Cost	65.0	216.6
Allowance for Funds Used During Construction	1.2	4.1
Total Plant Investment	66.2	220.7
Preproduction Costs	2.3	7.7
Inventory Capital	0.5	1.7
Cost of Construction Downtime	<u>21.0</u>	<u>70.2</u>
Total Capital Requirement	90.0	300.3

Thus, the total capital requirement for a S-H-U FGD retrofit for a 300-MWe power plant, equivalent in technical scope to the installation at Milliken Station, is approximately \$90 million (\$300/kW). However, when estimating costs for a particular installation, scope adjustments and site-specific factors, such as the scope of sorbent and gypsum-handling systems, sparing philosophy, production of a marketable by-product versus landfilling the gypsum, design coal properties, and design sulfur-removal efficiency, need to be taken into account.

IV.B.2 Operating Costs

Fixed operating and maintenance costs include operating labor, maintenance labor, administration and support, and maintenance materials to support the FGD facilities. Variable operating costs include sorbent, chemicals, utilities, waste disposal costs, and by-product credits. For a 300-MWe power plant with a 65-percent operating factor and other operating characteristics the same as in the preceding section, operating costs are shown in Table 4. Operating costs, including credit for gypsum sales, are \$4.6 million/year.

Table 4. S-H-U Operating Costs

Fixed O&M Costs	\$10⁶/yr
Operating Labor	0.60
Maintenance Labor	0.26
Maintenance Material	0.39
Administration/Support Labor	0.26
Fixed Costs	1.51
Variable Operating Costs	
Limestone	0.98
Formic Acid	0.08
Electric Power	2.30
Waste Disposal	0.09
Gypsum Sales	(0.34)
Variable Costs	3.11
Total O&M Cost	4.62

IV.B.3 Cost of Sulfur-Dioxide Removal

Estimated economics for the S-H-U process installed on a 300-MWe unit with a 65-percent capacity factor firing 3.2 wt% sulfur coal are listed in Table 5.

Table 5. S-H-U Costs for Sulfur-Dioxide Removal

Total Process Capital, \$/kW	179	
Total Capital Requirement, \$/kW	300	
SO ₂ Removal Efficiency, %	95	
	<u>Current \$</u>	<u>Constant \$</u>
Levelized Busbar Power Cost, mills/kWh	12.0	9.2
Levelized Cost, \$/ton SO ₂ recovered		
Capital Charge	377	292
Fixed O&M	51	40
Variable O&M	105	81
Total	533	413

Although these costs appear to be higher than other limestone wet flue gas scrubbing processes, economics is not the only factor that will promote the acceptance of the S-H-U process. A major factor is the highly beneficial environmental impact. Not only is the S-H-U process capable of SO₂ removals up to 98 percent, but instead of creating acres of sludge ponds, it produces a high-quality gypsum by-product, which wallboard manufacturers prefer over natural gypsum. Also, although not yet fully perfected, the process appears capable of producing a salable calcium-chloride brine concentrate, thus eliminating another disposal problem. Thus, the S-H-U process has the potential to operate essentially discharge free. In addition, the S-H-U process offers the potential for a small footprint by being located beneath the stack. Although this adds capital costs, it may be a preferred option for space-constrained plants.

V Conclusions

The objective of this project was to demonstrate high levels of NO_x and SO₂ control with minimum loss of efficiency while essentially eliminating waste-disposal problems. For the most part, this objective was achieved. Specific conclusions are:

- At full boiler load (145-150 MWe) and 3.0- to 3.5-percent economizer O₂, the LNCFSTM III system lowered NO_x emissions from a baseline value of 0.64 to 0.39 lb/MBtu, a 39-percent reduction. At a reduced boiler load of 80-90 MWe and 4.3- to 5.0-percent economizer O₂, the LNCFSTM III system lowered NO_x emissions from a baseline value of 0.58 to 0.41 lb/MBtu, a 29-percent reduction.
- Although measured boiler efficiency with the LNCFSTM III was about 1-percent lower than the baseline, operating conditions were not identical. When adjustments are made to bring both results to the same basis, boiler efficiency is essentially the same. Unburned carbon in fly ash was maintained below 4 percent, thereby maintaining fly ash salability.
- The S-H-U process is capable of removing up to 98 percent of the SO₂, as was demonstrated for short periods during this demonstration, but the unit must be designed properly to achieve this result. Many variables affect the actual sulfur-removal efficiency, including coal sulfur level, gas velocity and L/G ratio in the scrubber, pH of the slurry, and limestone properties and grind size. The effects of these variables were not studied as extensively as desired because the test program was shortened.
- Wallboard-grade gypsum can be produced and sold as a by-product rather than being landfilled. This significantly decreases waste-disposal problems.
- Technical problems prevented satisfactory operation of the brine-concentration system. Although progress was made in solving these problems, a salable concentrated brine solution was not produced as part of the MCCTD Project. However, it seems likely that with some additional effort, this system could be made to operate satisfactorily, thereby eliminating another waste-disposal problem.
- Some problems were initially encountered with the heat pipe air preheater in that the efficiency was not as high as expected. The problem was traced to contaminated heat-transfer medium in the tubes. Once the heat-transfer medium was replaced, the heat pipe air preheater worked as expected. Proper quality control by the manufacturer should prevent this problem from recurring. Although there was no air leakage between the air and flue gas, some air leakage did occur through the soot-blower penetrations. This problem should also be solvable.

- Total capital requirement for a 300-MWe power plant feeding 3.2 wt% sulfur coal with 95-percent sulfur-recovery efficiency is estimated as \$90 million (\$300/kW). For a 65-percent operating factor, levelized operating costs are estimated at \$533/ton of SO₂ removed (12.0 mills/kWh) on a current dollar basis and \$413/ ton of SO₂ removed (9.2 mills/kWh) on a constant dollar basis.
- The success of the MCCTD Project is verified by the continued use at Milliken station of the S-H-U process and the continued sale to a wallboard manufacturer of essentially all the gypsum produced.
- A detailed market analysis by NYSEG indicates a significant potential market for sales of the S-H-U process in the U.S. Through 2030, potential sales are estimated to be 5,700 MWe of retrofit capacity and 96,200 MWe of new capacity. The probability of meeting this market potential is considerably enhanced by the success of the MCCTD Project.

VI Abbreviations

CAAA	Clean Air Act Amendments
CaCl ₂	calcium chloride
CAPCIS	Corrosion and Protection Centre Industrial Services
CCT	clean coal technology
CEM	Continuous Emissions Monitor
CO	carbon monoxide
CO ₂	carbon dioxide
CONSOL	Consolidation Coal Company
E&HO	engineering and home office
EPRI	Electric Power Research Institute
ESP	electrostatic precipitator
FGD	flue gas desulfurization
FRP	fiberglass reinforced plastic
G/C	Gilbert/Commonwealth
I&C	instruments and controls
ID	induced draft
lb/MBtu	pounds per million Btu
L/G	Liquid-to-gas ratio
LNCFS TM III	Low-NO _x Concentric Firing System Level III
MCCTD	Milliken Clean Coal Technology Demonstration
Mills/kWh	millage per kilowatt hour
MWe	Megawatt
NAAQS	National Ambient Air Quality Standards
NO _x	nitrogen oxides
NSPS	New Source Performance Standards
NYSDEC	New York State Department of Environmental Conservation
NYSEG	New York State Gas & Electric Corporation
NYSERDA	New York State Energy Research and Development Authority
O ₂	oxygen
O&M	operation and maintenance
PEOA TM	Plant Emission Output Advisor,
PLC	programmable logic controller
PLC	programmable logic controller
Q Pipe	a heat pipe that preheats air
S-H-U	Saarberg-Hölter-Umwelttechnik
SNCR	selective noncatalytic reduction
SO ₂	sulfur dioxide
SOW	Statement of work
SPDES	State Pollution Discharge Elimination System
TSP	total suspended particles

VII References

CONSOL, 1996, *Detailed Test Plan for the Design Sulfur Coal Test of the S-H-U Scrubber at the NYSEG Milliken Station*, (prepared for NYSEG), May 10, 1996

CONSOL, 1996, *Unit 2 Electrostatic Precipitator Performance Test Results, Before and After Modification*, (prepared for NYSEG), December 1996.

CONSOL, 1999a, *Milliken Clean Coal Technology Demonstration Project—Data Report for 1.6 percent Sulfur Coal Tests of the S-H-U Scrubber at Milliken*, (prepared for NYSEG), May 1999.

CONSOL, 1999b, *Milliken Clean Coal Technology Demonstration Project—Data Report of the Design Coal Tests of the S-H-U Scrubber at Milliken*, prepared for NYSEG), July 1999.

CONSOL, 2000, *Milliken Clean Coal Technology Demonstration Project—Data Report of the High-Sulfur Coal Tests of the S-H-U Scrubber at Milliken*, (prepared for NYSEG), January 2000.

Electric Power Research Institute, 1991, *Economic Evaluation of FGD Systems*, EPRI GS-7193.

Janati, K., 1997, *Program Results from a Comprehensive Assessment of Chemical Emissions from New York State Electric & Gas Corporation's Milliken Station Unit 2, Lansing, NY—Final Report*, (prepared by Carnot for NYSEG), July 1997.

New York State Electric & Gas Corporation, 1993, *Milliken Clean Coal Technology Demonstration Project—Public Design Report*, November 1993.

New York State Electric & Gas Corporation, 1999, *Milliken Clean Coal Demonstration Project—Project Performance and Economics Report—Final Report*, Volumes I and II, April 1999.

U.S. Department of Energy, 1992, *Comprehensive Report to Congress—Clean Coal Technology Demonstration Project—Milliken Clean Coal Technology Demonstration Project*, September 1992.

VIII Bibliography

Baron, E.S., III, Gauffillet, G., and Jackson, C.E., 1995, "Milliken Station Demonstration Project FGD Retrofit Update—1995," Fourth Annual Clean Coal Technology Conference (Denver, CO), September.

Brigham Young University, 1997, "Report on Modeling of Combustion Gases—Unit 2 Boiler," (prepared for NYSEG), September.

CONSOL, 1994a, "Unit 1 LNCFS Level 3 and Unit 2 Baseline Test Program Results," (prepared for NYSEG), August.

CONSOL, 1994b, "Unit 2 Electrostatic Precipitator Performance Test Results, Baseline Report," (prepared for NYSEG), August 1994.

CONSOL, 1995, "Heat Pipe Performance and Acceptance Testing," (prepared for NYSEG), June.

CONSOL, 1996a, "Unit 2 Electrostatic Precipitator Test Results, Performance Upgrade Tests," (prepared for NYSEG), October.

CONSOL, "Unit 1 LNCFS Level 3 and Unit 2 Performance Upgrade Tests," (prepared for NYSEG), October.

CONSOL, 1997a, "Heat Pipe Performance Test Results—Interim Report No. 2," (prepared for NYSEG), February.

CONSOL, 1997b, "Heat Pipe Performance Report Test Results—Interim Report No. 3," (prepared for NYSEG), June.

CONSOL, 1997c, "The SHU Low Sulfur Flue Gas Desulfurization Summary Report at Milliken Station," (prepared for NYSEG), August.

CONSOL, 1997d, "Comparison of ESPert™ Model Predictions with Unit 2 Electrostatic Precipitator Performance—Interim Report," (prepared for NYSEG), November.

CONSOL, 1998a, "Inspection and Demolition of Stebbins Tile Test Module," (prepared for NYSEG), February.

CONSOL, 1998b, "Heat Pipe Performance—Final Report," (prepared for NYSEG), July.

DeVito, M.S., Oda, R.L., and Marker, B.L., 1995, "Milliken Clean Coal Technology Demonstration Project, Unit 2 Electrostatic Precipitator Upgrade Performance Testing—Interim Report to NYSEG," November.

DHR Technologies, Inc., 1995a, “Plant Economic Optimization Advisor Commercialization Plan (PEOA),” (prepared for NYSEG), March.

DHR Technologies, Inc., 1995b, “Plant Economic Optimization Advisor (PEOA) Test Plan,” (prepared for NYSEG), April.

Ellis, C.M., Savichky, W.J., and Elia, G.G., 1993, “The Milliken Station Clean Coal Demonstration Project: There’s More to It Than Concrete and Steel,” Second Annual Clean Coal Technology Conference (Atlanta, GA), September.

Harvilla, J., Burns, S., Savichky, W., Mahlmeister, M., and Watts, J., 1998, “Milliken Clean Coal Technology Demonstration Project,” Sixth Clean Coal Technology Conference, May.

Mahlmeister, M.E., Statnick, R.M., Glamser, J., Jackson, C.E., Hoffman, J.E., Aliasso, R.E., and Elia, G.G., 1992, “Overview of the Milliken Station Clean Coal Demonstration Project,” First Annual Clean Coal Technology Conference, November.

Mereb, J.B., Statnick, R.M., and Chang, S.S., 1996, “Milliken Clean Coal Technology Demonstration Project—Unit 1 LNCFS Level 3 and Unit 2 Baseline Test Program Results—Final Report,” December.

O’Dea, D.T., Jackson, C.E., and Elia, G.G., 1994, “Milliken Station Demonstration Project: FGD Retrofit Update,” Third Annual Clean Coal Technology Conference (Chicago, IL), September.

Savichky, W.J., O’Dea, D.T., and Watts, J.U., 1997, “Anatomy of an Upgraded Pulverized Coal Facility: Combustion Modification Through Flue Gas Scrubbing,” Fifth Annual Clean Coal Technology Conference (Tampa, FL), January.

Southern Research Institute, 1997, “Droplet Testing at the New York State Electric & Gas Milliken Station,” (prepared for NYSEG), January.

Stouffer, M.R., McCoy, D.C., Theodore, F.W., and Yoon, H., 1992, “New York State Electric & Gas Gypsum Demonstration Project: Results of High Sulfur Test Center Pilot Testing—Report to NYSEG,” November.

U.S. Department of Energy, 1999, “Advanced Technologies for the Control of Sulfur Dioxide Emissions from Coal-Fired Boilers,” Clean Coal Technology Topical Report Number 12, June.

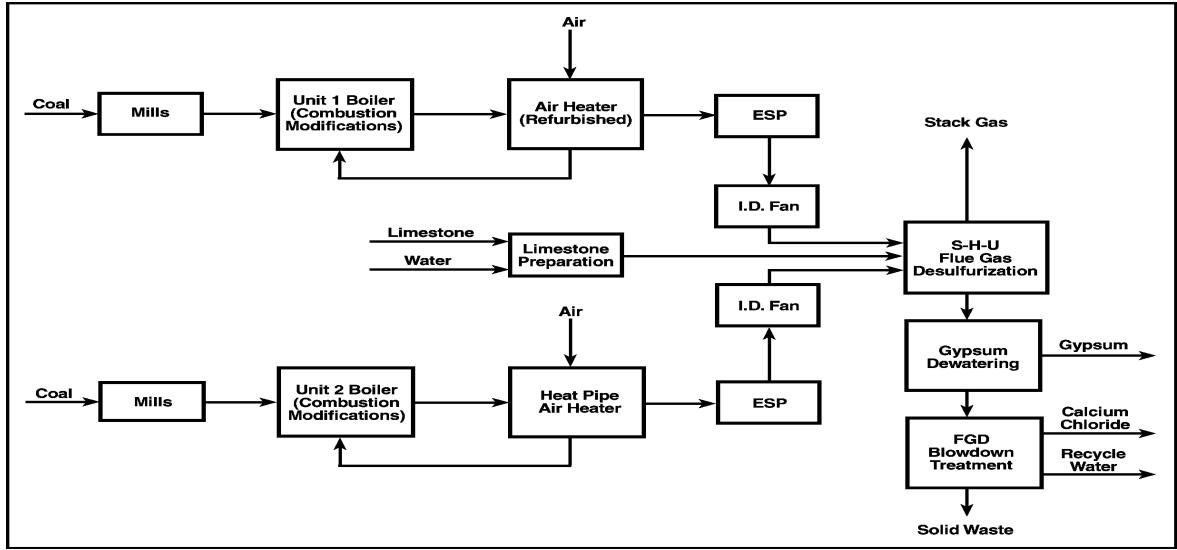


Figure 1. Block Diagram for the MCCTD Project

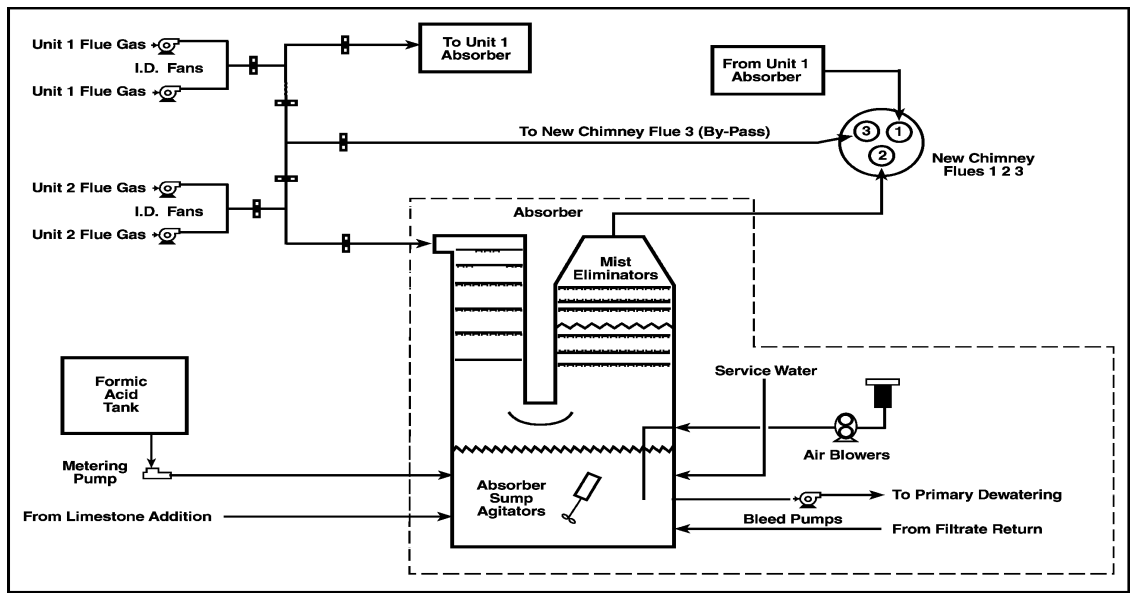


Figure 2. S-H-U Flow Diagram

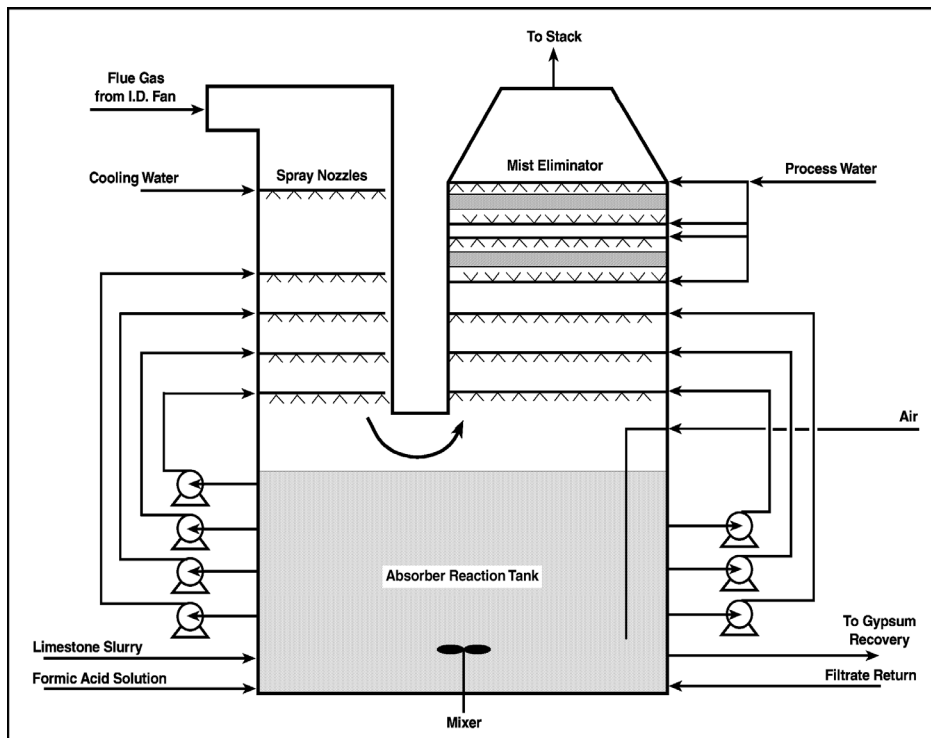


Figure 3. S-H-U FGD Absorber

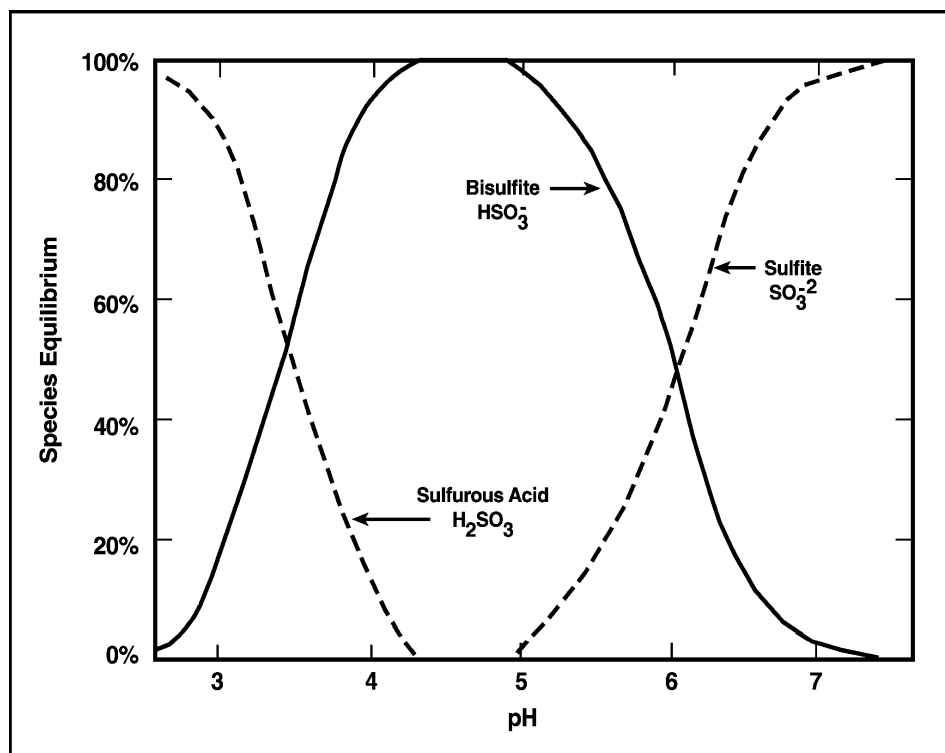


Figure 4. Equilibrium Diagram for the $\text{H}_2\text{SO}_3 - \text{HSO}_3^- - \text{SO}_3^{2-}$ Systems

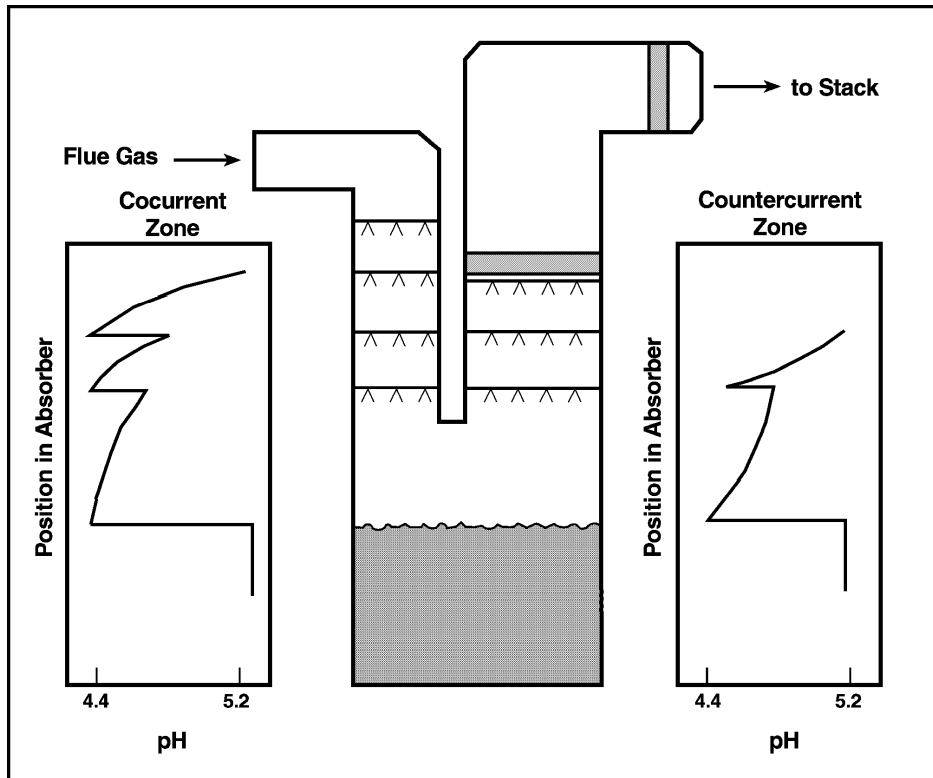


Figure 5. pH as a Function of Position in S-H-U Cocurrent/Countercurrent Absorber

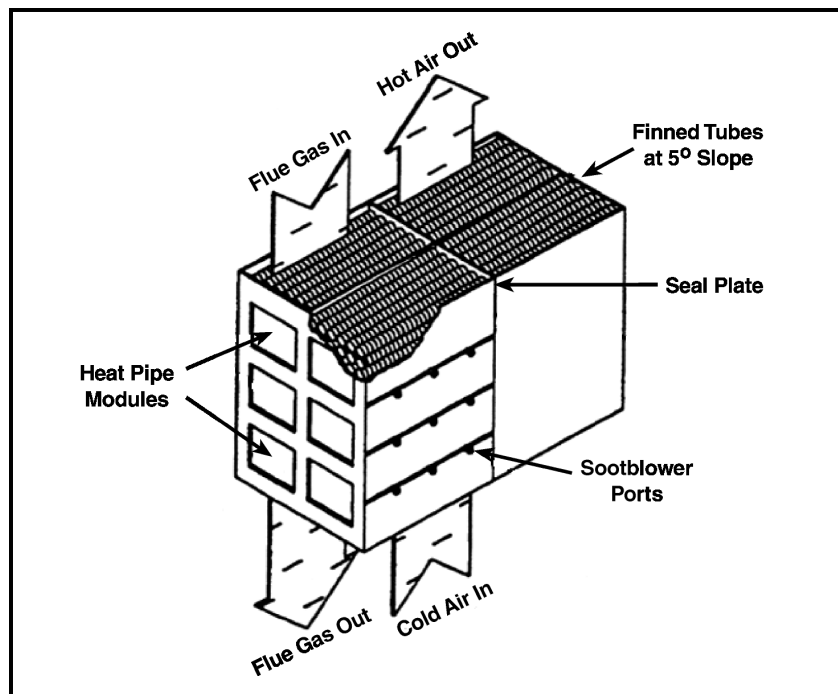


Figure 6a. Typical Heat Pipe Air Heater Construction

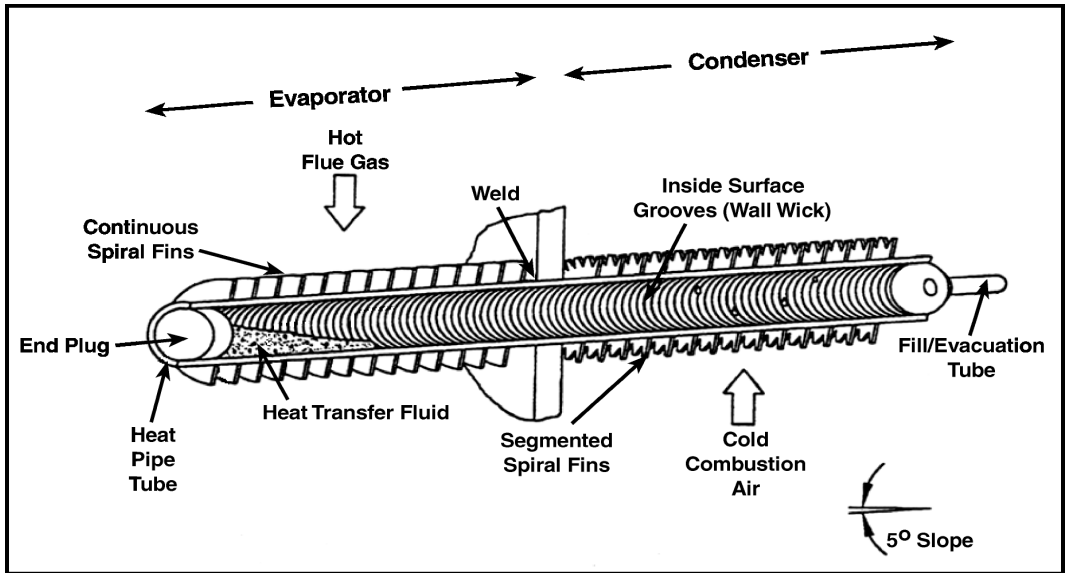


Figure 6b. Heat Pipe Cross Section

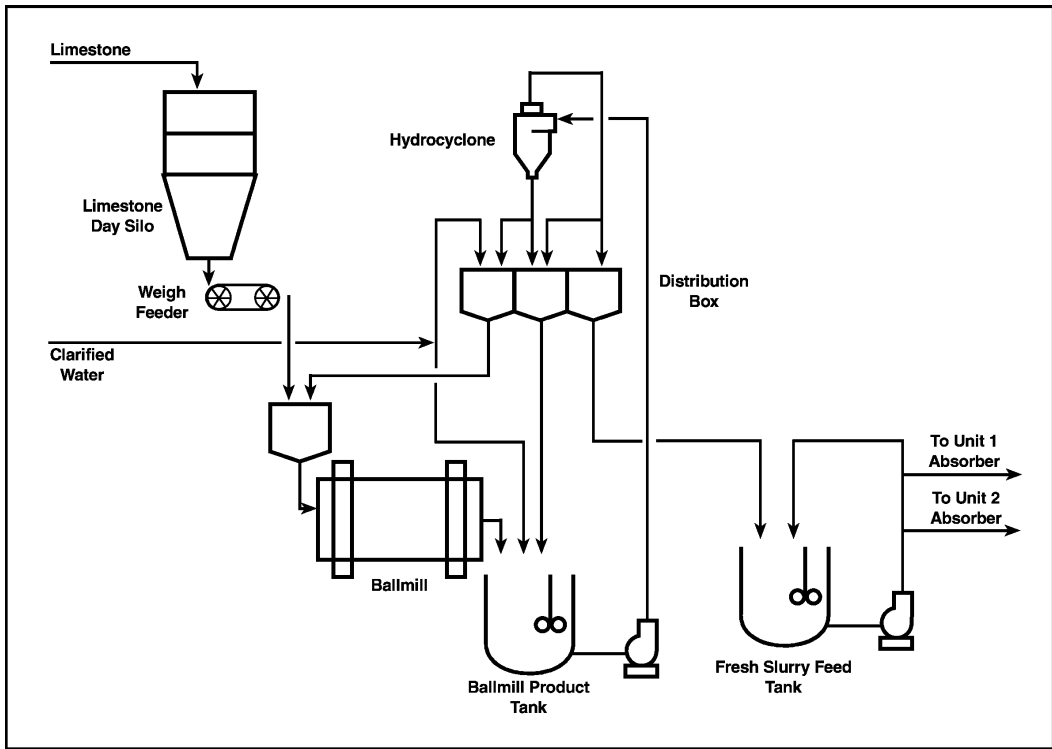


Figure 7. Limestone Preparation Flow Diagram

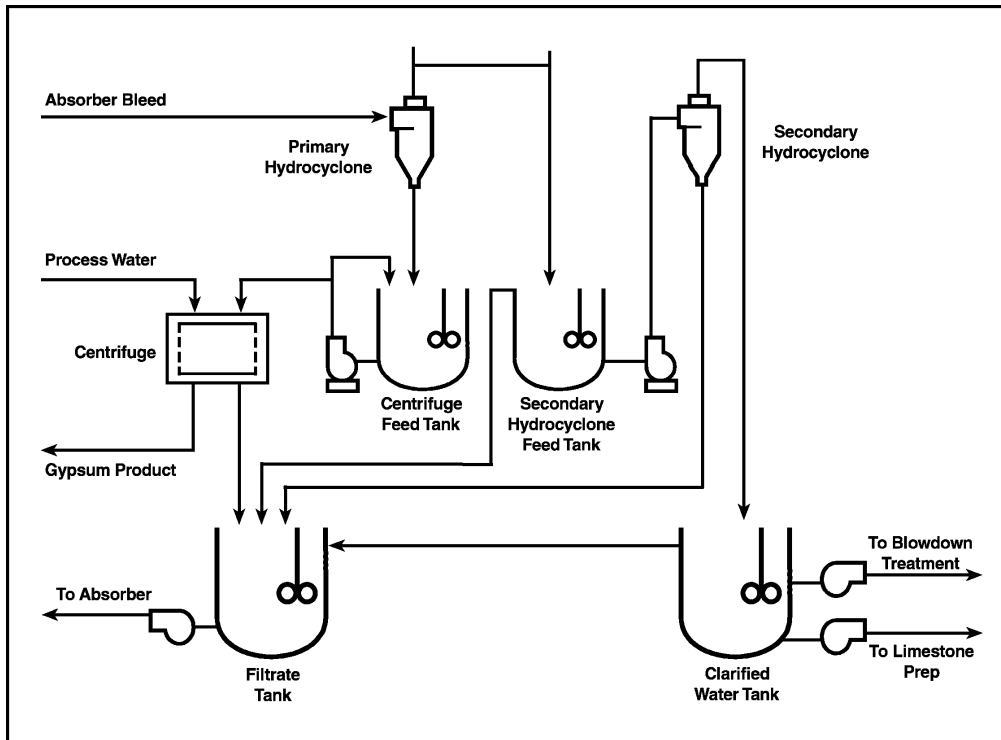


Figure 8. By-Product DeWatering Flow Diagram

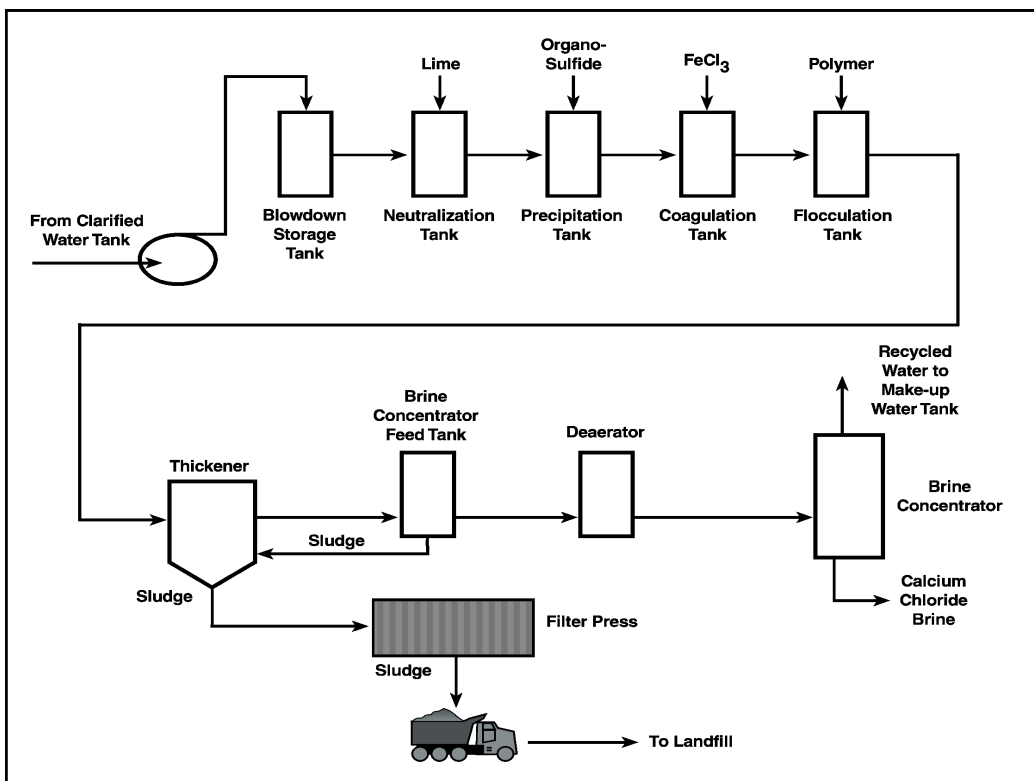


Figure 9. Blowdown Treatment/Brine Concentration Flow Diagram

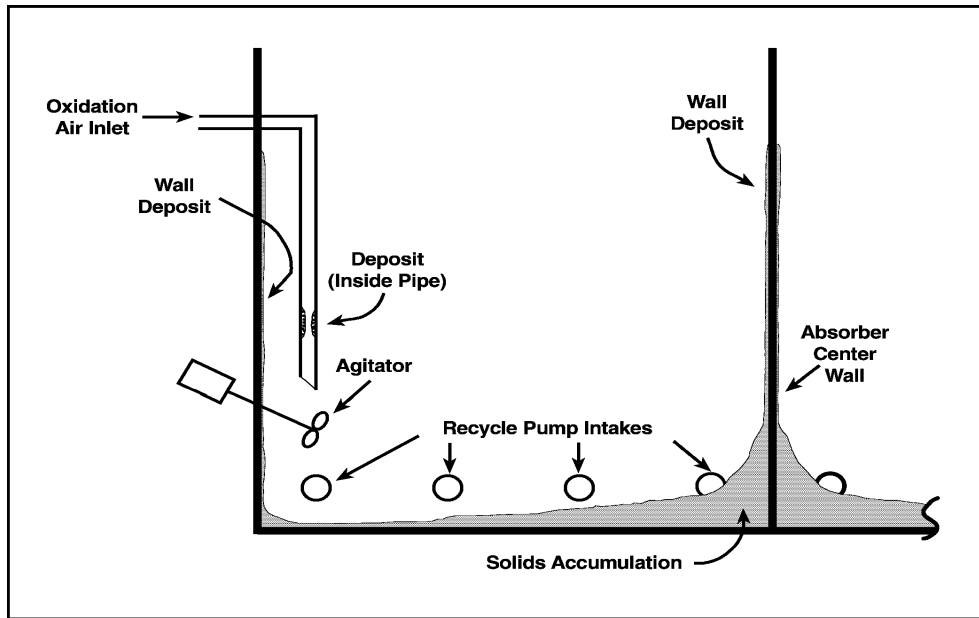


Figure 10. Schematic Showing Approximate Locations of Solid Deposits in Absorber

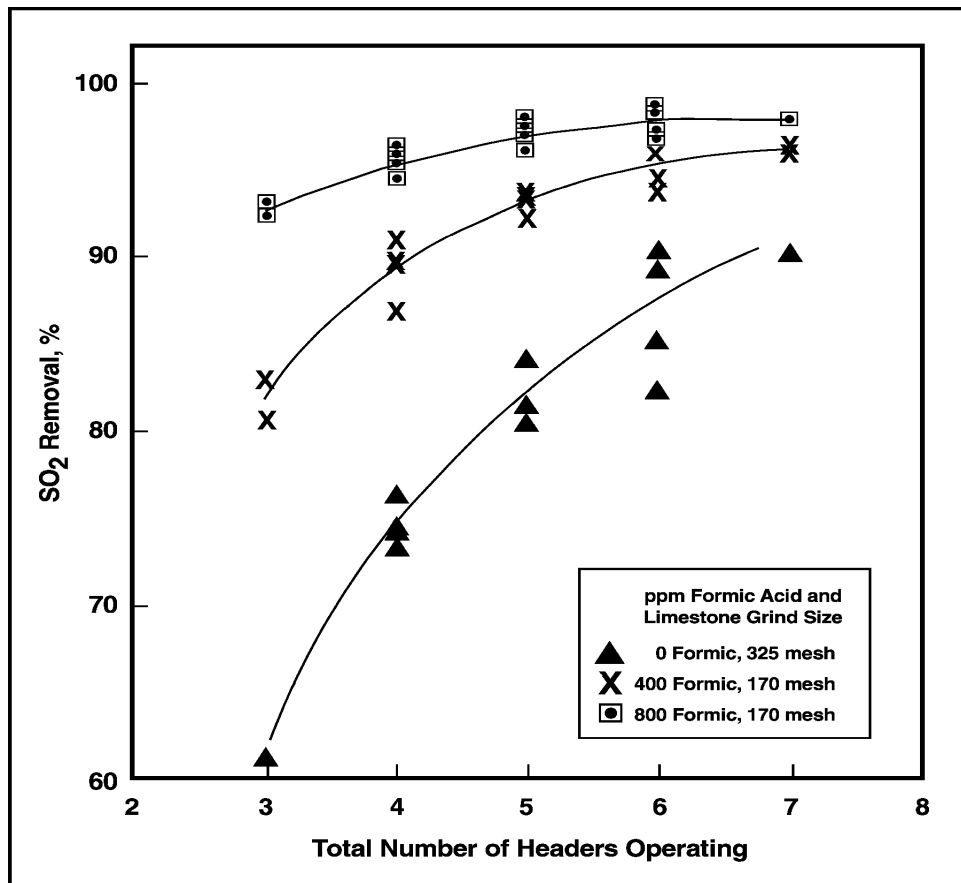


Figure 11. SO₂ Removal vs. Total Number of Headers in Operation

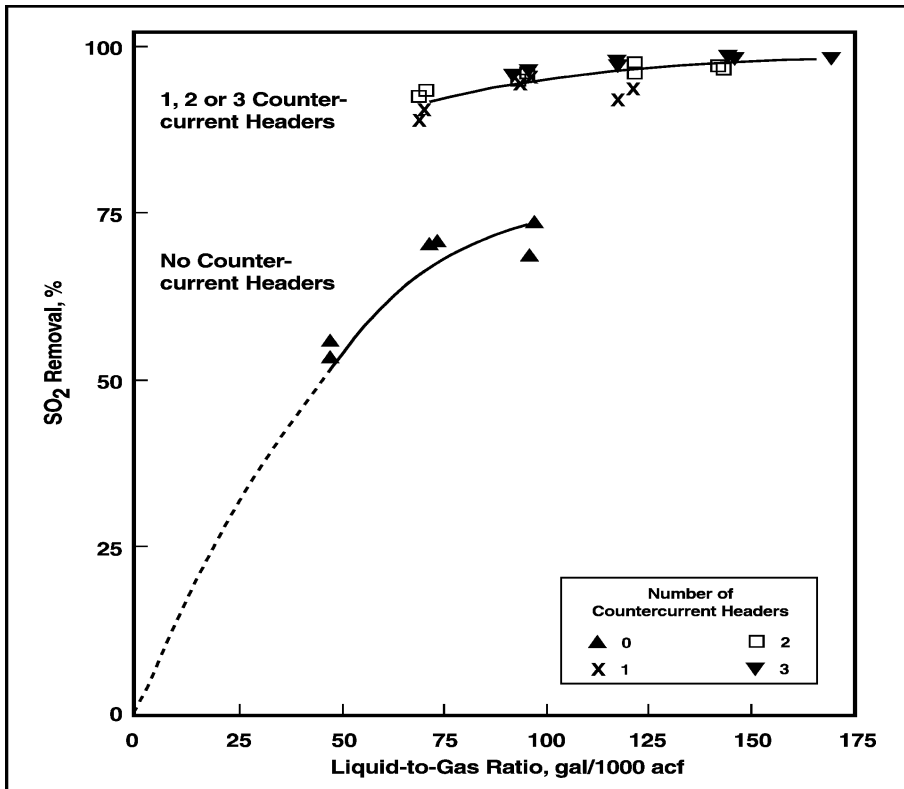


Figure 12. SO₂ Removal vs. Liquid/Gas Ratio

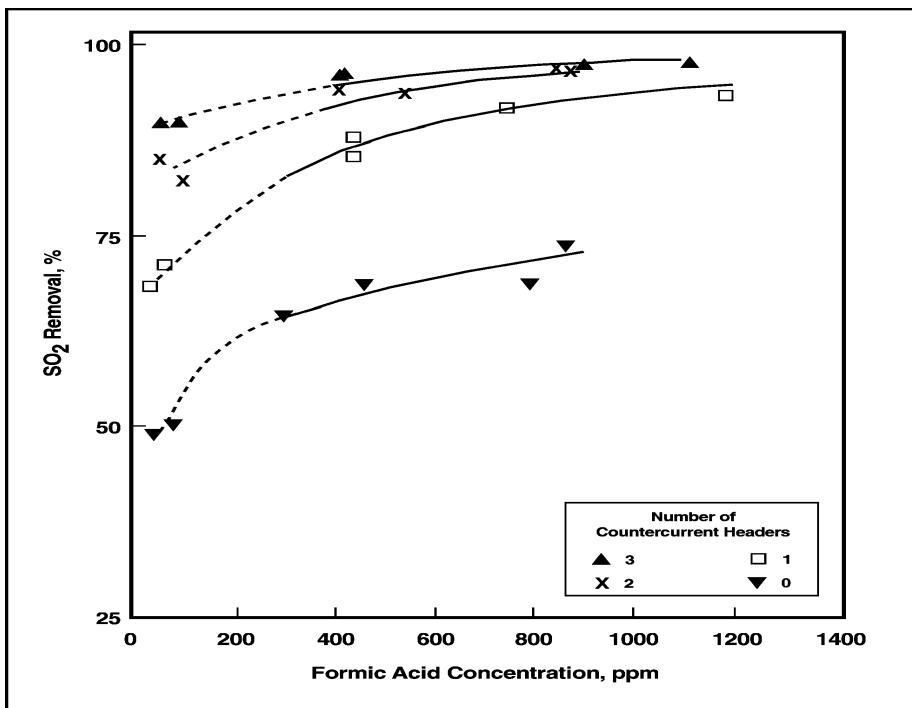


Figure 13. SO₂ Removal vs. Formic Acid Concentration

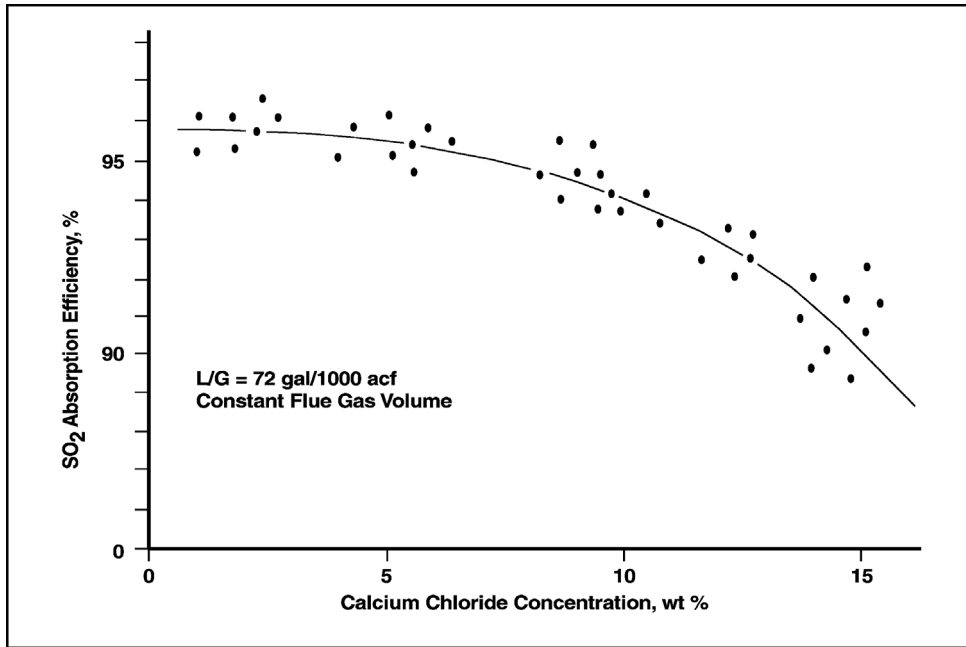


Figure 14. Effect of CaCl₂ Concentration on SO₂ Absorption Efficiency

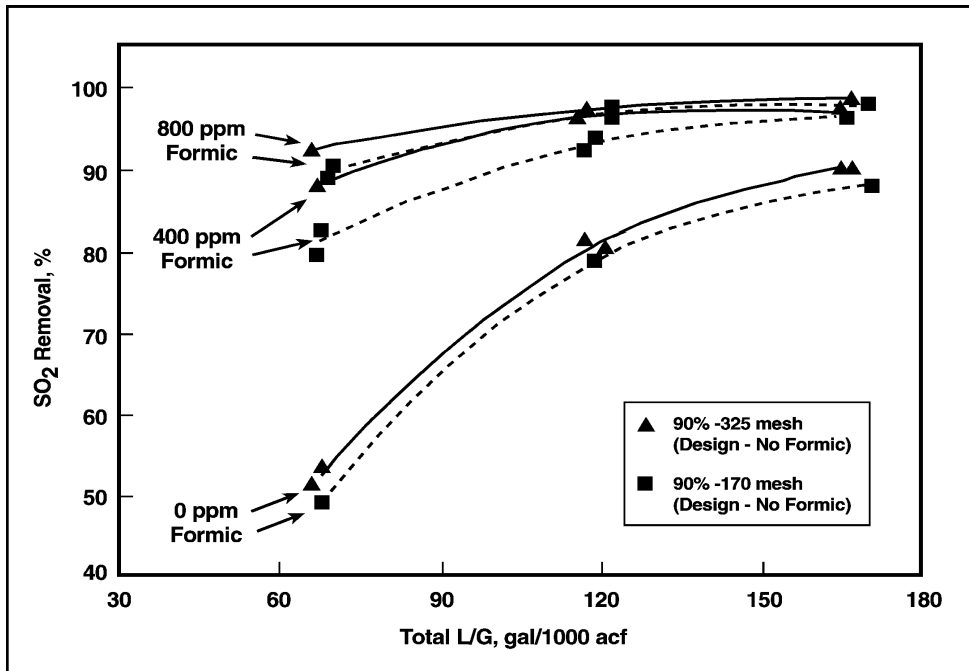


Figure 15. Effect of Limestone Grind on SO₂ Removal